Chapter 3

The atmospheric and vacuum crude distillation units

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The distilling of petroleum products from crude oil to some extent or other has long been practiced. Certainly the ancient Egyptians, Greeks, and Romans had some form of extracting a flammable oil from, probably, weathered crude oil seepage. It wasn't though until the turn of the nineteenth and twentieth century that crude oil well drilling was first discovered and commercialized. Originally the crude oil was refined to produce essentially kerosene (lamp oil), and a form of gasoline known then as Benzine (as opposed to Benzene already being produced from coal) and the residue used as pitch for calkin and sealing. The lamp oil or kerosene was produced to provide a means of illumination, later a lighter cut known as naphtha was produced for the same purpose but used in special pressurized lamps.

The production of these early distillates was made by cascading the crude oil through successive stills each operating at successively higher temperatures. This is shown in the following diagram Figure 3.1.

The crude enters the first still to be heated to a temperature that vaporizes the light components. The residue from this still enters the second one and heated to a higher temperature to vaporize the Benzine fraction. The residue enters the third still and heated to remove the Kerosene fraction. The residue from this still is a very light fuel oil which may be further heated and partially vaporized to give a pitch of sorts as a residue and a distillate which could be used as fuel. This distillate would later become the Diesel or Gas Oil fraction and used in the developing diesel engine.

The vapor from each still passes through a small packed wash section before being condensed and collected in a condensate drum. A portion of the condensate is returned to the top of the wash section as the wash liquid, similar to the reflux stream in modern distillation towers. Usually steam was injected through the liquid phase of each still to facilitate vaporization and to strip out the light ends.

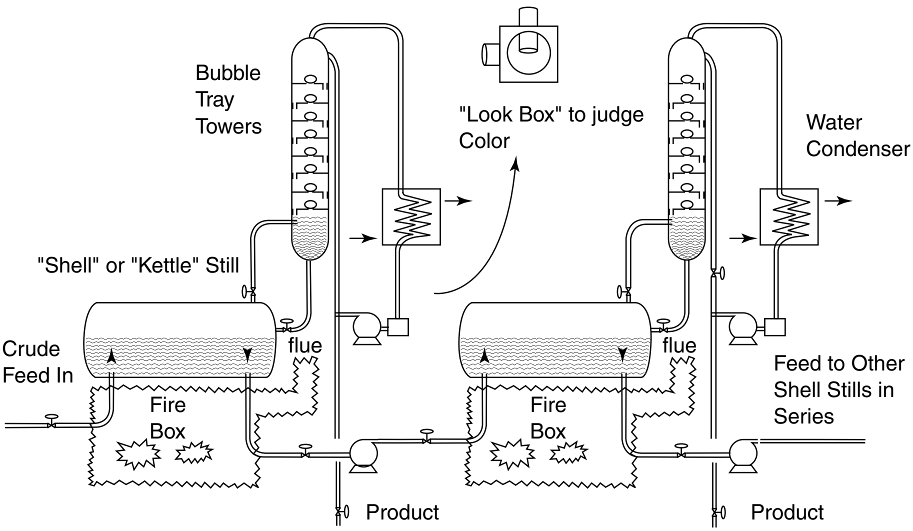


Figure 3.1. Early continuous pipestill schematic.

This type of crude distillation was superseded by the continuous fractionation tower used in modern petroleum refining. This type of crude 'Still' remained in operation well after the second World War in some oil refineries. The author recalls such a still in operation as late as 1956 in the refinery that he was employed at, although there were also in operation three other modern crude oil fractionating units. Indeed one major oil company even today call the crude distillation unit a 'Pipestill'. This chapter now continues with the description of a present day atmospheric and vacuum crude oil distillation units. It will be divided into two parts:

Part 1 The Atmospheric Crude Distillation Unit (CDU)

Part 2 The Vacuum Crude Distillation Unit (VDU)

Both parts will contain a process description with a schematic flow sheet, a discussion on the development of the material balance, a description and discussion on the design characteristics of the units, and finally a worked design example.

3.1 The atmospheric crude distillation unit

Process description

The first process encountered in any conventional Refinery is the Atmospheric Crude Distillation Unit. In this unit the crude oil is distilled to produce distillate streams

which will be the basic streams for the refinery product slate. These streams will either be subject to further treating down stream or become feed stock for conversion units that may be in the Refinery Configuration. A schematic flow diagram of an atmospheric crude unit is shown in Figure 3.2.

Crude oil is pumped from storage to be heated by exchange against hot overhead and product side streams in the Crude Unit. At a preheat temperature of about 200–250°F water is injected into the crude to dissolve salt that is usually present. The mixture enters a desalter drum usually containing an electrostatic precipitator. The salt water contained in the crude is separated by means of this electrostatic precipitation. The water phase from the drum is sent to a sour water stripper to be cleaned before disposal to the oily water sewer.

It must be understood however that this 'de-salting' does not remove the organic chlorides which may be present in the feed. This will be discussed later when dealing with the tower's overhead system.

The crude oil leaves the desalter drum and enters a surge drum. Some of the light ends and any entrained water are flashed off in this drum and routed directly to the distillation tower flash zone (they do not pass through to the heater). The crude distillation booster pump takes suction from this drum and delivers the desalted crude under flow control to the fired heater via the remaining heat exchange train.

On leaving heat exchanger train, the crude oil is heated in a fired heater to a temperature that will vaporize the distillate products in the crude tower. Some additional heat is added to the crude to vaporize about 5% more than required for the distillate streams. This is called over flash and is used to ensure good reflux streams in the tower. The heated crude enters the fractionation tower in a lower section called the flash zone.

The unvaporized portion of the crude leaves the bottom of the tower via a steam stripper section, while the distillate vapors move up the tower counter current to a cooler liquid reflux stream. Heat and mass transfer take place on the fractionating trays contained in this section of the tower above the flash zone. Distillate products are removed from selected trays (draw-off trays) in this sections of the tower. These streams are stream stripped and sent to storage. The full naphtha vapor is allowed to leave the top of the tower to be condensed and collected in the overhead drum. A portion of this stream is returned as reflux while the remainder is delivered to the light end processes for stabilizing and further distillation.

The side stream distillates shown in the diagram are:

- Heavy gas oil (has the highest Boiling Point)
- Light gas oil (will become Diesel)
- Kerosene (will become Jet Fuel)

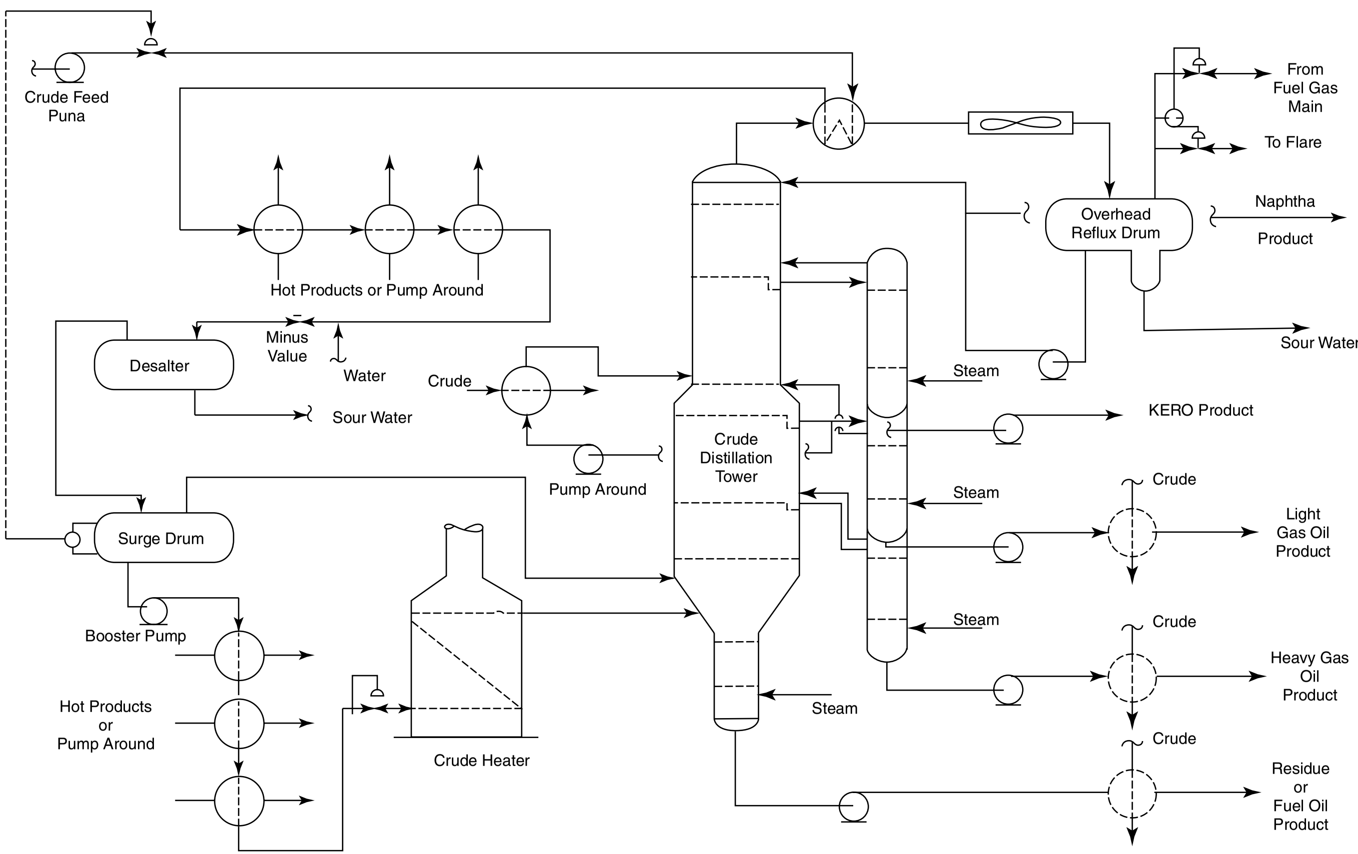


Figure 3.2. A typical atmospheric crude distillation unit.

A 'Pump around' section is included at the light gas oil draw off. This is simply an internal condenser which takes heat out of that section of the tower. This in turn ensures a continued reflux stream flow below that section. The product side streams are stripped free of entrained light ends in separate stripping towers. These towers also contain fractionation trays (usually four but sometimes as many as six) and the side stream drawn off the main tower enters the top tray of its respective stripper. Steam is injected below the bottom tray and moves up the tower to leave at the top, together with the light ends strip out, and is returned to the main fractionator at a point directly above the side stream draw-off tray. These side stream stripper towers are usually stacked one above the other in a single column in such a way as to allow free flow from the side stream draw-off tray to its stripper tower. On a few occasions, where the particular side stream specification requires it, the stripping may be effected by reboiling instead of using steam. One such requirement maybe in the Kero side stream if this stream is to be routed directly into jet fuel blending and therefore must be dry.

The residue (unvaporized portion of the crude) leaves the flash zone to flow over four stripping trays counter current to the flow of stripping steam. This stripping steam enters the tower below the bottom stripping tray. Its purpose primarily is to strip the residue free of entrained light ends. The fact that this steam enters the flash zone it also enhances the flashing of the crude in this zone by creating a reduced partial pressure for the liquid/vapor separation. This becomes an important factor in the design and operation of the atmospheric crude distillation unit. The stripped residue leaves the bottom of the unit to be routed either through the unit's heat exchanger system and the to product storage or hot to some down stream processing unit such as a vacuum distillation unit or a thermal cracker.

The development of the material balance for the atmospheric crude distillation unit

The knowledge of the material balance in any refining process is important both for ensuring its proper design and later for its proper operation. Because of the relative number of streams involved this is particularly so in the case of the atmospheric crude distillation unit. The operation of this unit also is critical to the performance of down stream units such as catalytic crackers and catalytic reformers. The material balance for any specific operation required of the unit, (for example, maximizing naphtha feed to the catalytic reformer), from a particular crude feed source also sets the performance parameters of the unit. This includes the amount of reflux to be generated, at which section this reflux is to be generated, distillate draw-off temperatures, tower overhead conditions, flash zone conditions and the like.

Whole crude TBP curve and assay data

The development of the unit's material balance begins with an 'in depth' examination and analysis of the crude oil feed's assay. The first step in doing this activity is to

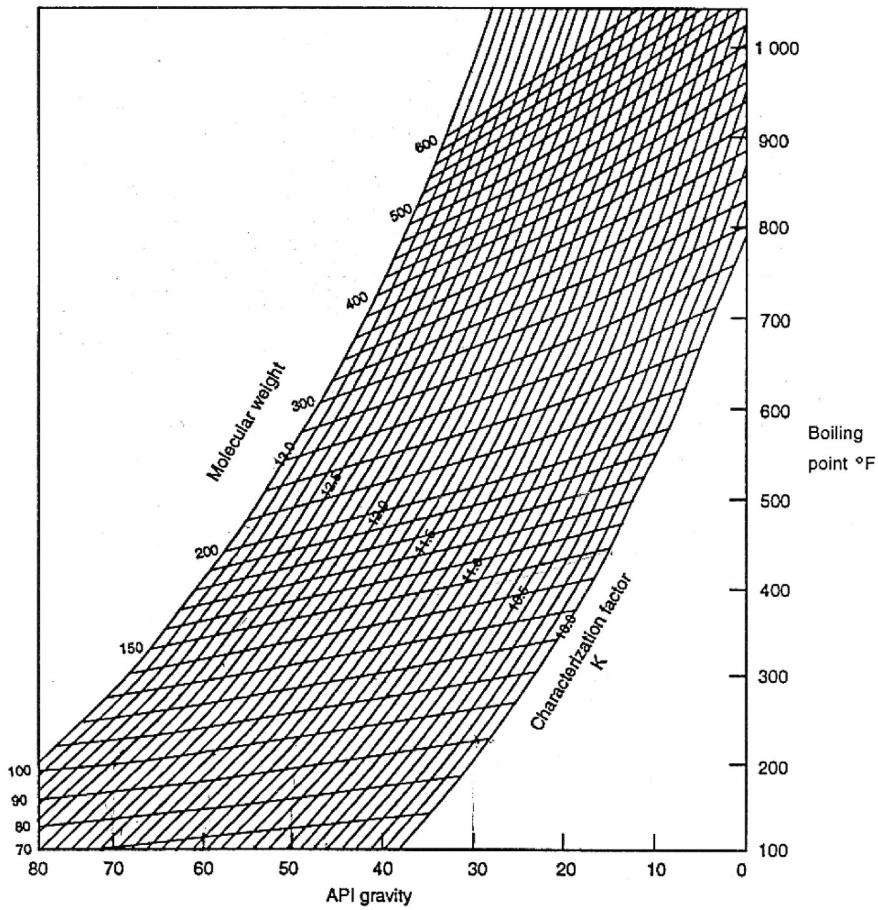


Figure 3.3. Boiling points V's molecular weight and gravity (°API).

break up the crude feed's TBP distillation curve into its pseudo components and to assign the product characteristics relative to each component's mid boiling point or mid volume point on crude.

The properties of each pseudo component are provided either by the assay data giving the gravity versus mid volume percent (or mid boiling point), and by the application of Figure 3.3. This chart gives the component's molecular weight based on boiling point (i.e., mid boiling point) and its gravity in °API. Every care must be made to establish these properties as accurately as possible. It will form the basis of any design process or definitive study work that may be carried out on this unit.

Developing the side stream product TBPs

The first step is to divide the crude feed's TBP curve into the product yield (in percent volume) and in terms of that product's temperature cut range. This starts with establishing the cut point of the residue. That is the temperature on the TBP curve at which it is intended to separate the total distillate from the residue. Then the distillate portion of the TBP is divided into its product cuts such as Gas Oils, Kerosene, and the total overhead Naphtha, or some other combination of products. The volume of these cuts and temperature ranges represent the yield of each product that will be produced as a percentage of the whole crude.

From the cut points of each of the products their ASTM curves are developed using the method described in Chapter 1. The initial boiling point (IBP) of each cut (except the overhead Naphtha) is fixed by the fractionation capability of the distillation unit. This term 'fractionation capability' for a crude oil distillation unit is measured as the difference in temperature between the 95% volume point of the lighter cut's ASTM distillation and the 5% volume point of the adjacent heavier cut. This difference may be positive (Gap) or negative (Overlap). A gap indicates good separation while the overlap indicates poor separation. The ability to separate the fractions efficiently decreases as the products become heavier. Thus, one can expect an ASTM gap between the overhead product and the first side stream to be around 25°F, while that between a third and fourth side stream to have an ASTM gap around −10°F (an overlap of 10°F).

The side stream TBP curves are now developed using these concepts and following these stepwise procedures.

Step 1. Establish the cuts and cut ranges on the crude TBP that represents the products that will be produced in this unit. For example—The overhead product will be a full range naphtha and will contain all the gas in the crude and the distillate to a cut point of 400°F. The first side stream will be a Kerosene cut beginning at 400 and ending at 500°F on the crude TBP curve. The next side stream will be a gas oil boiling between 500 and 650°F on the TBP curve. These will have yields as % volume on crude from the TBP curve. As an example typical cuts are given as Figure 3.4.
From this figure:

Naphtha (gas to 400°F) = 30 %vol on crude.
Kerosene (400–500°F) = 10.5 %vol on crude.
Gas oil (500–650°F) = 14.5 %vol on crude.
Residue (+650) = 45.0 %vol on crude.

Step 2. Predict the ASTM end point and the ASTM 90% point using the figures given in Chapter 1 of this Handbook for the full range Naphtha (i.e., the overhead product).

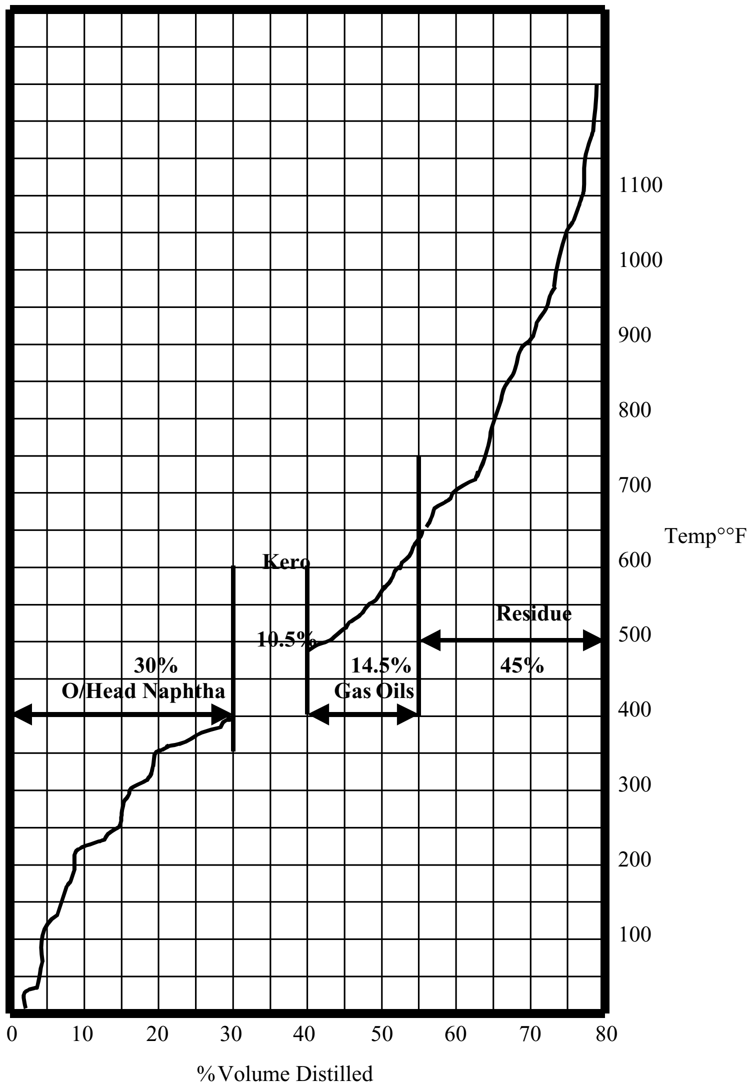


Figure 3.4. TBP and product cut points for a typical mid east crude.

Step 3. Plot these on the probability chart, also given in Chapter 1 and draw the straight line through the 30 %vol point. This will be sufficient to determine a meaningful ASTM curve from which a TBP curve can be produced.

Step 4. The ASTM curves for the remaining side stream products are developed using the curves in Chapter 1 for fixing their ASTM end points. The 5% points on both

ASTM curves are then fixed using a reasonable ASTM Gap or overlap between the products thus:

Step 5. The 5% ASTM point for the kerosene cut will be the 95% point of the full range naphtha plus 25°F gap. Similarly the 5% ASTM point for the gas oil will be the 95% point of the kerosene plus a 0°F gap.

Step 6. The ASTM curves for the side streams are drawn as straight lines on the probability chart between their respective end points and their predicted 5% points.

Step 7. Convert the developed ASTM curves to TBP curves using the Edmister correlations and as described in Chapter 1.

Step 8. Extend the front end of the full naphtha to include the gas portion and the light distillate below the 30 %vol point. Step this section off in mid boiling points to simulate real hydrocarbon components such as C5's, C6, C7, etc. This will become important later in establishing the reflux drum pressure and temperature.

An example of this method is given later in the worked example of an Atmospheric Crude Distillation Unit process design.

Developing the product volume, mass, mol balance

Using the component breakdown (pseudo components) and the product TBP curves, calculate each product volume rate, mass rate, and mole rates using the following steps:

Step 1. Establish the crude feed flow rate in terms of volume (usually BPSD), then calculate its mass flow (say in lbs/hour) and molal flow, using the crude feed breakdown table described in the previous section.

Step 2. Develop each products' specific gravity using its component composition and each component's specific gravity as given in the Crude Breakdown Table.

Step 3. Develop each product's mol weight similar to step 2 and again referencing the Crude Breakdown Table. There is relationship between gravity, boiling point, and mole weight. This is given in Figure 3.3.

Step 4. From the data developed in steps 2 and 3 calculate the quantity of each product in terms of BPSD, lbs/hr, moles/hr. The sum of each of these product quantities must equal the quantity of the crude oil feed calculated in Step 1.

This completes the description of the material balance development.

The design characteristics of an atmospheric crude distillation fractionating tower

In modern day refining the separation of the basic products from the crude feed is generally accomplished in a single atmospheric distillation tower. There are circumstances however that lead to the use of two towers to accomplish this. These

circumstances usually occur where there is an abnormally high quantity of light components in the feed (such as crude spiked with naphtha) or where the unit capacity revamp calls for such a configuration. In these circumstances a light cut is flashed off and fractionated in a 'Primary' tower. The bottoms from this tower is then heated and vaporized to enter the flash zone of a 'Secondary' tower where it is fractionated to meet the conventional overhead, and side distillate products.

This section of this chapter will deal only with the single atmospheric crude distillation process however.

The tower flash zone

As described earlier the crude oil feed is heated to its required temperature for separating its distillate products from the unvaporized portion (i.e., the residue) by heat exchange with products being cooled or condensed and finally by a fired heater. This heated crude enters a section of the atmospheric fractionating tower at a point below the fractionating trays. This section of the tower is called The Flash Zone. The temperature and pressure conditions in this zone are such as to allow the correct amount of vapor to be flashed from the heated crude to provide the quantities of the distillate product streams and the internal reflux to control their quality.

These flash zone conditions of temperature and pressure are determined taking into consideration the shape of the crude oil's equilibrium flash vaporization (EFV) curve, the pressure (usually the partial pressure in the case of the atmospheric tower) existing in this section of the tower and the temperature on the EFV corresponding to the volume % cut of the total vapor to be flashed.

Setting the flash zone partial pressure

Because steam is used to strip the residue leaving the tower free of light ends this steam enters the flash zone from the top stripping tray. The steam influences the pressure condition at which the hydrocarbon vapors separate from the residue. This pressure condition is the partial pressure exerted by the steam. It is calculated by the equation as follows:

$$\text{Partial pressure of HC vapor} = \frac{\text{moles HC vapor}}{\text{moles HC vapor} + \text{moles steam}} \times \text{Total pressure}$$

The total pressure of the flash zone is determined by the pressure in the tower's overhead reflux drum plus the pressure drop over the overhead condenser(s), and the

total distillate trays and internals above the flash zone. For an estimate of the flash zone partial pressure the following rules of thumb may be considered:

Overhead reflux drum pressure	5 psig
Pressure drop over the condensers	7 psi
Pressure drop over trays etc	10 psi (assumes 40 trays @ 0.25 psi per tray)
Assume ΔP over internals are negligible	
Then total pressure	22 psig.

Assume that the steam from the residue stripper will be 1.2 lbs/hot gallon of residue.

Setting the flash zone temperature

Calculate the EFV of the whole crude using the method described in Chapter 1 of this Handbook. This EFV curve is at atmospheric pressure. The temperature of the flash at atmospheric pressure is read off as the % volume vaporized to meet the amount of distillate products *and* the over flash required by the process. This over flash is usually fixed at between 3 and 5% volume on crude. Its purpose in the process is to provide that extra heat in order to generate sufficient reflux down flow over the trays to satisfy the prescribed degree of separation between the products.

This atmospheric flash temperature is now adjusted to the temperature at the previously calculated partial pressure existing in the flash zone. This may be done by reading the temperature at the partial pressure using the PVT curve shown as Figure 1.A.1 in the appendix to this chapter. This is the flash zone temperature.

Other features of the flash zone section

Most crude oils contain sulfur in some form or other and organic chlorides which are not removed in the pre-treating desalter. Corrosion from these impurities is particularly virulent in mixed phases (liquid/vapor) and at the elevated temperature experienced in the unit's flash zone, and made more so by the presence of steam. A cladding of 11/13 chrome is usually applied as a protective cover over the tower's carbon shell in this section. This cladding should be extended also to include the residue stripper and the section of the tower containing the first 4–6 wash trays above the flash zone. These trays and the residue stripping trays should also be of 11/13 chrome alloy.

Effective separation of the distillate vapor phase and the unvaporized residue phase is enhanced by the inclusion of a 'swirl' at the inlet of the flash zone. The crude from the fired heater is routed through this 'swirl' which is an inverted trough and extends around 2/3rds of the tower circumference. The swirling action caused by this forced flow of the mixed phases allows the lighter vapor phase to separate to a large extent from the heavier liquid phase. The final separation occurs on the top stripping tray.

The fractionator overhead system

The fractionator overhead equipment has four functions. These are:

- To condense the overhead vapors including the stripping steam
- To return and control the reflux condensate to the tower
- To collect and dispatch the overhead product
- To separate and dispose of the condensed steam

These functions are accomplished in several different ways. Some of these overhead configurations are described in the following paragraphs.

Configuration 1

This is probably the most common system and is shown in Figure 3.5.

The total overhead vapor leaves the top of the tower at its dew point. It first enters a shell and tube exchanger where it is partially condensed. The cooling medium is cold crude oil feed from storage. The partially condensed overhead leaves the shell side of this crude oil exchanger to be further and totally condensed by either a trim water cooler or, more usually an air condenser. The condensate from this final condenser enters an overhead condensate drum. This drum is designed to allow the complete separation of the condensed steam in the overheads from the hydrocarbon condensate. This condensed aqueous phase is collected in a 'boot' located below the main condensate drum and is pumped under level control to the refinery's sour water disposal system. A portion of the hydrocarbon phase is pumped to back to the fractionating tower to enter above the top tray. This stream is usually flow controlled which is reset by a tower top temperature control. The stream enters the tower through a spray system designed to ensure good distribution over the top tray.

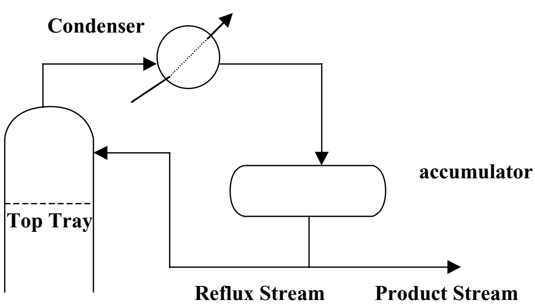


Figure 3.5. The most common overhead configuration.

The remaining hydrocarbon condensate leaves the drum under level control and is pumped to storage or a downstream process plant. The operating temperature and pressure of the condensate drum is vital to the operation of the fractionating tower. These conditions set the parameters for flash zone operation and intermediate product draw-off temperatures. These conditions are set and operated as follows.

The drum pressure. This is maintained by two pressure control valves operating on a split flow control. One of the valves is connected to the refinery flare while the other is connected to the unit's fuel gas supply. When the pressure in the drum exceeds the pressure set point of the controller, the first valve opens to the flare thus reducing the drum pressure. When the pressure falls below the set point then fuel gas is introduced by the second valve to correct this condition.

The drum temperature. The drum temperature is held at the bubble point of the condensate or slightly super cooled. This temperature is controlled either by fan pitch or by louvres on the final air condenser. Should the final condenser be a shell and tube water trim exchanger then a bypass of the condensate around the exchanger activated by a temperature/flow control is installed. Such an arrangement can also be used on an air condenser.

The tower overhead temperature. The tower overhead temperature in this configuration is the dew point of the overhead product vapor at the partial pressure of the hydrocarbon.

Note: The partial pressure for this case is the total moles of hydrocarbons (includes the Reflux moles) divide by the total moles of vapor (includes the steam present) times the total overhead pressure.

The tower overhead pressure. The overhead tower pressure in this case is the pressure of the overhead condensate drum plus the pressure drops of the heat exchangers, and the associated overhead condensate piping. This pressure drop may be taken as 7–9 psi.

Configuration 2

This second configuration is usually installed on units of high capacity at throughputs above 70,000 BPSD. Its purpose is to maintain the size of the overhead equipment to manageable dimensions. This includes drums and the heat exchangers associated with the system (Figure 3.6).

In this configuration the total overhead vapors are again condensed by two exchangers as described in Configuration 1. In this second case, however, the condensate from the first condenser is collected in a drum and returned to the tower as reflux. The vapor from this first drum includes the overhead product hydrocarbon and the uncondensed

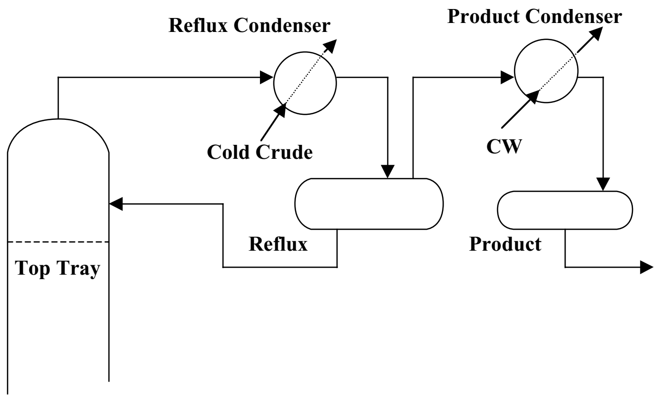


Figure 3.6. The 'Two Drum' configuration.

steam from the fractionator. This vapor is condensed in the second overhead condenser (again usually an air condenser) to be collected in a second condensate drum as the overhead distillate product and water. Disposal of the separated aqueous and hydrocarbon streams follow the same system as in Configuration number 1. In this configuration the reflux drum acts as an additional theoretical tray because the reflux liquid is in phase equilibrium with the product vapor.

The product drum pressure and temperature. This again is the controlling pressure for the fractionator as a whole. It is set at the pressure which allows the vapors to be condensed at a reasonable temperature at or below its bubble point. In moderate climates with average air temperatures of 60°F this temperature will be around 90–100°F with a pressure of 5 psig. Control of the drum pressure and temperature remains the same as that for configuration 1.

The reflux drum temperature. The conditions in this drum may be considered as those for the top tray of the tower in Configuration 1. The temperature therefore is the dew point of the product distillate at the partial pressure of the hydrocarbons vapors leaving the drum. This temperature is usually controlled by a vapor bypass over the exchanger on a flow reset by temperature control valve. Alternatively it may be controlled by the crude oil being bypassed on a similar flow/temperature reset basis. The partial pressure in this case is the moles of product vapor only divided by the total moles (including the steam) times the total absolute pressure. This total pressure is the pressure in the distillate product drum (say 5 psig) plus the pressure drop of the air condenser and piping (say about 3 psi).

The tower overhead temperature and pressure. The tower top conditions in the case of this second configuration is calculated as the dew point of the reflux stream entering

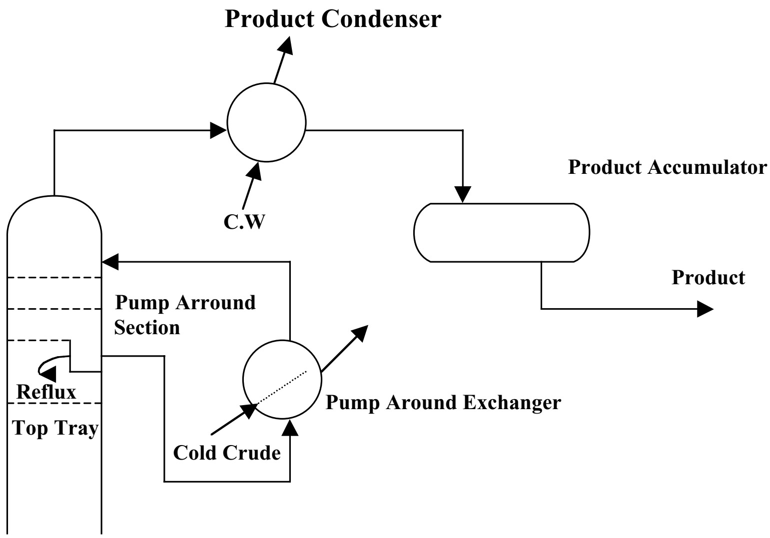


Figure 3.7. The tower top pumparound configuration.

the tower from the reflux drum. This temperature will be at the partial pressure of the total hydrocarbons leaving the tower.

Configuration 3

In this configuration the Tower itself is used to generate the internal reflux from the overhead product (Figure 3.7).

In this configuration the reflux is generated by a pump around system taking the liquid from the top section of the tower, and cooling it by heat exchange with cold crude oil. This cooled stream is then returned to the top tray of the tower. This is a conventional pump around system. About three trays are used for this pump around section in the tower. The tower top temperature remains as the dew point of the overhead product distillate at its partial pressure with steam. The temperature of the liquid leaving the tower as the pump around will be between 5 and 7°F higher than the tower top temperature. The internal pump around section may be assumed to be equivalent to one half an actual tray for mass transfer.

Discussion

All three of these configurations have their advantages and disadvantages. These are discussed briefly in the following paragraphs:

Configuration 1. As said earlier this is the most common configuration. It is relatively easy to control with really only two set points that one has to be concerned with. They are the drum pressure and the tower overhead temperature. There is good heat transfer between the condensing vapors and the crude oil. The overall heat transfer coefficient is high at around 70–80 Btu/hr/sqft/°F and advantage is taken in the LMTD of the latent heat of the vapors being transferred to the low temperature of the incoming crude. The biggest disadvantage with this system is that the crude to the first condenser is of course at a much higher pressure than the vapors condensing on the shell side. The rupture of a tube in the exchanger bundle or a leak in the tube plate will cause crude to contaminate the overhead product and 'dirty' the tower overhead trays. This can be quite a disaster if the downstream units (particularly the Cat Reformer) is fed directly from the condensate drum without some intermediate surge drum or tankage.

Configuration 2. This has the advantage of providing a very high LMTD in condensing with heat exchange against crude. This is due to the higher temperature now experienced in the vapors leaving the tower. As in the case of configuration 1, the overall heat transfer coefficient in this case is also high. There still remains though the problem of contamination but in this case downstream units have a measure of protection in that only the vapors from the reflux drum are condensed to make the product. The top trays of the tower however will be contaminated as before. The two drum system's biggest advantage however is more aligned to the installed cost of the unit. Moderately sized equipment are much cheaper than large items which may have to be field fabricated. Cost savings may in any case be achieved in the transportation costs of the drums and heat exchangers.

Configuration 3. The only major advantage in this configuration is that the liquid pumparound can be circulated at a higher pressure than the incoming crude. Thus, the problem associated with tube fracture or leakage contaminating the tower and product is eliminated. There is also an advantage in better control of the tower top temperature by the flow control of the pumparound itself. This too can be reset by temperature, but as an ex plant operator the author is not particularly in favor of temperature reset as a finite means of control. There is no other apparent advantage to this system. Indeed the inclusion of additional trays in the tower for the pumparound is quite expensive. The trays would need to be of monel and some monel lining of the shell will be required in this section as protection against chloride attack.

Some design considerations

Corrosion

The top section of the atmospheric distillation tower is very prone to corrosion from HCl. To combat this corrosion and to protect the carbon steel shell from damage monel is used in tray manufacture and as lining for the tower shell. This material is also used to line the first condenser shell interior and the tubes may also be fabricated

in monel. Ammonia either in its anhydrous form or as a solution, is injected into the tower top and at specific intervals into the top section tray spaces. The purpose of the ammonia is to neutralize as far as possible the HCl. The injection of the ammonia is controlled by the Ph reading of the condensate water leaving the boot of the product condensate drum.

Corrosion is most virulent at the liquid interface on trays. Consequently in most modern fractionating towers monel lining of the shell is usually applied as strips on the top four or six trays. These strips installed at the interface of the liquid/vapor on these trays the strips will be about 4–6 ins wide. The injection of anhydrous ammonia in the case of configuration number 2 does cause problems in the formation of ammonium chloride crystals occurring in the ‘dry’ section of the condenser system. That is the section where the hydrocarbon is condensing but the steam is still in its vapor phase. The injection of aqueous ammonia solution instead of the anhydrous form solves this problem.

Sulfur compounds are also a problem with respect to corrosion particularly in the ‘cold’ liquid phase of the distillate product drum. The drum shell in this case is gunnite lined. This lining is applied to the bottom section of the main drum and extended to include the ‘boot’ in which the condensed water is concentrated. Gunnite is best described as a cement lining and is applied in the same way as refractory lining, by plastering it on to the shell.

The distillate drum

The design of the distillate drum includes space for the complete separation of the aqueous phase (condensed steam) from the hydrocarbon phase. It is important for the hydrocarbons to be as free of water as possible to protect the downstream units. The performance of these units such as hydro-treaters and, more particularly the catalytic reformer are badly effected by the presence of water in their feed streams.

The design of the drum to allow this separation is based on one of three laws. These laws are expressed by the following equations:

$$\text{Stoke's Law: } V = 8.3 \times 10^5 \times \frac{(d^2 \Delta S)}{\mu}$$

(used when Re number is < 2.0)

$$\text{Intermediate Law: } V = 1.04 \times 10^4 \times \frac{d^{1.14} \Delta S^{0.71}}{S^{0.29}\text{c} \times \mu^{0.43}}$$

(used when the Re number is 2–500)

$$\text{Newton's Law: } V = 2.05 \times 10^3 \times \{(d\Delta S) \div \text{Sc}\}^{1/2}$$

(used when the Re number is > 500)

where

Re number $= \frac{10.7 \times dv\mathrm{Sc}}{\mu}$

V = Settling rate in ins per minute.
d = Droplet diameter in inches.
S = Droplet specific gravity.
$S\mathrm{c}$ = Continuous phase specific gravity.
ΔS = Specific gravity differential between the two phases.
μ = Viscosity of the continuous phase in Cps.

The following criteria may be used as a guide to estimating the droplet size:

Lighter phase	Heavy phase	Minimum droplet size
0.850 Sc and lighter	Water	0.008 inches
Heavier than 0.850	Water	0.005 inches

Piping

Vapor piping from the tower top should be sized for a minimum economic pressure drop. As a guide vapor rundown lines should be sized to meet an average pressure drop of 0.2 psi per 100 ft and should never exceed 0.5 psi/100 ft. The vapor (and mixed phase) piping and the drums in the tower overhead system should be so arrange as to allow free drainage towards the distillate drum (or in the case of configuration 2 to each successive drum). The vapor lines should never be pocketed.

Distillate drums should be located at least 15 ft above the center line of the respective distillate/reflux and the sour water pumps (pumps are normally located at grade). For safety this measurement should be taken from the bottom tangent line of the vessel. Piping to the pump should be sized for a 0.25 psi/100 ft pressure drop, while the discharge piping from the pump may be reduced to meet a pressure drop of 2.5–4.0 psi/100 ft.

The side streams and intermediate reflux sections

There are generally two or three side streams in the atmospheric crude distillation unit. There may be one more but a total of three side streams is the norm. The unit, that is going to be developed as an example in this work, has three side streams. These are:

- Kerosene
- Light gas oil
- Heavy gas oil.

With three side streams there will be usually two pumparound sections. The purpose of these pumparound sections is to create the proper internal reflux to that section of the tower below the location of the respective pumparound. The pumparound takes suction from a draw-off tray in the tower. Usually this tray is also a product draw off. The pump delivers the pumparound stream through a heat exchanger where it is cooled before returning it to a location in the tower about two to three trays above the draw-off tray. This cooling liquid flowing down the tower to the draw-off tray acts as an internal condenser. The hot vapors rising up the tower through the draw-off tray contain the side stream product, some internal overflow and the vapor phase of all the lighter products and their overflow. This particular side stream product and overflow is selectively condensed by the cold pumparound. It enters the draw-off tray where the product quantity and pumparound leave the tower, and the remaining condensate overflows to the tray below the draw off. This over flow material is the internal reflux stream and its quantity determines the fractionation between this lighter side stream product and the heavier side stream product drawn from the tower at some point below.

The side streams

Side streams are withdrawn from selective and specially designed trays in the main tower. Each stream is usually combined with the respective pump around as described earlier. When withdrawn from the tower the product is in equilibrium with the total vapor rising through the draw-off tray. Consequently the product liquid will contain entrained light ends which will affect the required product properties such as flash point and its ASTM distillation. To correct this the side stream product is steam stripped in a separate trayed stripping column. This arrangement is shown in Figure 3.8.

The side stream from the main tower is routed to a distributor over the top tray of the stripping column under level control. The stripping column contains four stripping trays. Steam is introduced below the bottom tray and flows up the tower counter current to the liquid product. The light ends contained in the stripper feed are removed by the steam and both the stripout and steam leave the top of the stripper to enter the main tower at the vapor space directly above the draw-off tray.

The stripped product leaves the well at the bottom of the stripper to be pumped under flow control to storage or a downstream unit's surge drum. The level in the well of the stripper activates the level control valve of the incoming stripper feed.

The side stream draw-off temperature. In theory the side stream product draw off from the main column will be in equilibrium with the vapor rising from the tray. In other words it will be at its bubble temperature at the partial pressure of the hydrocarbons on the tray. However, this liquid phase will contain some entrained light ends (Thus the need for steam stripping) which will affect the actual temperature of the draw off. J. W. Packie in his published work on crude oil distillation developed a curve which

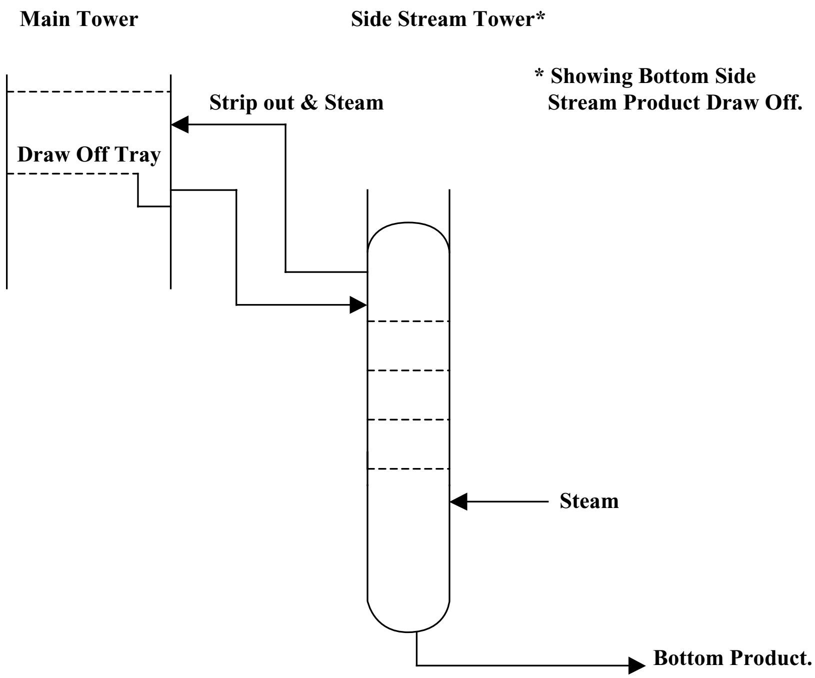


Figure 3.8. A sides tream stripper.

takes into consideration this difference between the theoretical draw-off temperature and the actual. This curve is given as Figure 3.9.

In arriving at the partial pressure in this calculation Packie assumes that all the vapor lighter than the draw-off is inert. With this as a parameter the draw-off temperature can be calculated using the following step wise process:

Step 1. Fix the amount of steam present in the vapor *passing through* the draw-off tray (*Note*: This does not include the steam returning from the particular draw-off stripper which enters the vapor space above the draw-off tray). The amount of stripping steam recommended is given in Figure 3.10.

Step 2. Fix the total pressure on the tray taking the estimated tower top pressure and each tray pressure drop above the draw-off tray (see the previous paragraphs on flash zone and overhead system). Estimate the position of the draw-off tray. As a guide the following rule of thumb can be used. With tray 1 as the top tray in the column:

Kero draw off	tray 12
Light gas oil	tray 22
Heavy gas oil	tray 32

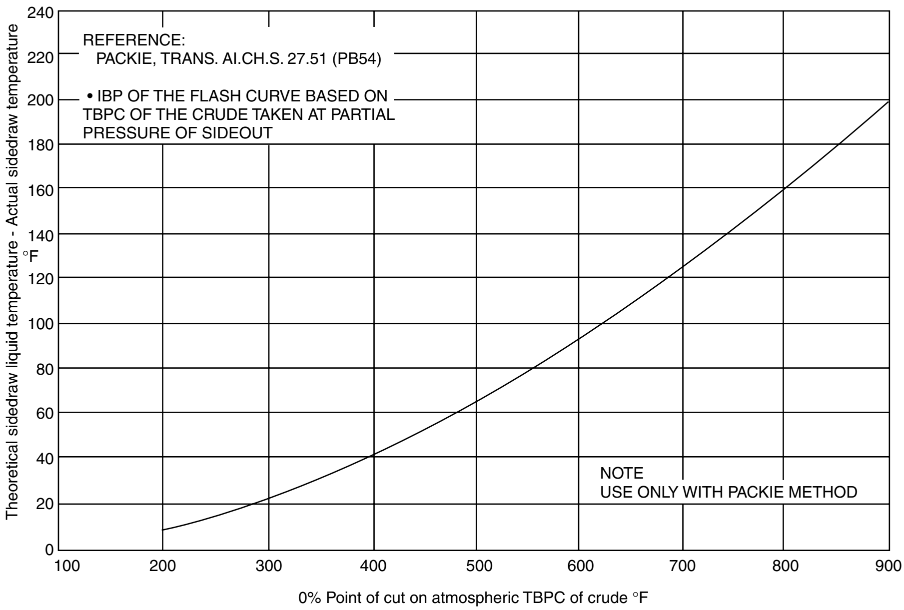


Figure 3.9. The difference between theoretical and actual draw-off temperatures.

There are usually 40 fractionating trays (i.e. Trays above the flash zone) and 4 bottom stripping trays in conventional distillation towers. These amounts will be checked later using the relationship of reflux to tray ratio for fractionation criteria.

Step 3. Calculate the FRL (Flash Reference Line) for the draw off TBP cut. In doing this use the TBP cut; not the TBP end points. This calculation has been described under the section on the Flash Zone. Establish its IBP at atmospheric pressure. This will be the 0 %vol temperature on the FRL.

Step 4. Predict the amount of overflow that will leave the draw-off tray as a liquid reflux to the tray below. Again this will be checked by the relationship of trays and reflux later. For this calculation the following rules of thumb for the ratio of moles overflow to moles product may be used:

Overflow from the HGO draw off 2.9
Overflow from the LGO draw off 1.2
Overflow from the Kero draw off 0.9 to 1.0

Step 5. Calculate the partial pressure on the draw-off tray as follows:

$$\text{Partial pressure} = \frac{(\text{total draw-off product vapor moles} + \text{total overflow moles})}{(\text{total hydrocarbon vapor moles} + \text{Steam passing through tray})} \times \text{total pressure}$$

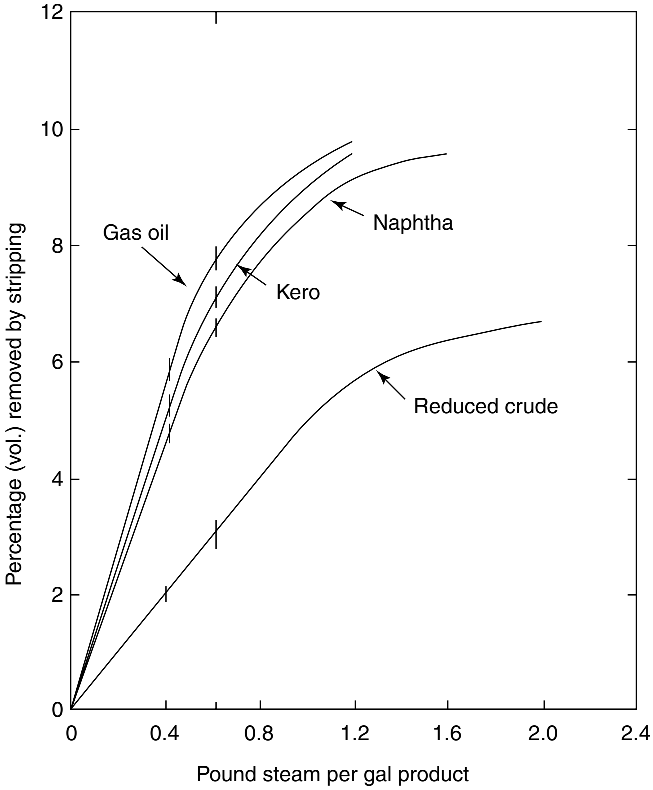


Figure 3.10. The amount of stripping steam.

Step 6. Using the vapor pressure curves (given in the appendix as Figure 1.A.1) Relate the IBP of the cut at atmospheric pressure (step 3) to the calculated partial pressure.

Step 7. The temperature arrived at in Step 6 is the theoretical temperature of a liquid in equilibrium with its associate vapor phase. In actual practice the fact that a vapor phase has been continually bubbling through the liquid on the tray, that liquid will contain entrained light ends from the vapor. The actual temperature will therefore be somewhat lower than the calculated, and this difference is provided by Figure 3.9. Establish the actual temperature of the side stream draw off. Note this will also be the temperature of the pumparound liquid also.

Side stream stripping. The side stream enters a steam stripper at the draw-off temperature. It is steam stripped and the amount of stripping steam and the corresponding strip out may be read off from Figure 3.10. The steam introduced below the bottom stripping tray is usually superheated at 50 psig. The temperature of the stripout and

steam leaving the top of the stripper to enter the main tower can be taken as 5°F below the temperature of the stripper feed. The mol weight of the stripout is taken as the average of the vapors lighter than the draw off. A heat balance can now be carried out over the stripper with an unknown being the stripped product enthalpy and temperature. The product outlet temperature and enthalpy is calculated. By solving the equation:

Heat in = Heat out.

Side stream strippers for each product draw off are usually stacked in one single column. It becomes an exercise in layout to locate these stripper sections. It begins with establishing the height of the bottom stripper section which will ensure sufficient head to provide available NPSH for the product rundown pump. Thereafter each successive section is located to ensure free flow of the liquid feed into the respective stripper, and to minimize the length of the vapor return line to the main tower.

The pump arounds

The mechanism of the pumparounds and their purpose has already been described and discussed. It remains now to examine how to evaluate and detail quantitatively their size in terms of the duty they will be required to perform. This will be followed with some criteria that need to be observed in their design.

Total pumparound duties. This begins by establishing the Tower bottom temperature (Residue outlet temperature) and then conducting an overall Tower heat balance. To calculate the Tower bottom temperature consider the following diagram (Figure 3.11):

And then the following calculation steps:

Step 1. Establish the material entering the top stripping tray. This will be the total residue product which now includes the overflash as a liquid, and the stripout material (also as liquid). This is taken as being at the flash zone temperature.

Step 2. Calculate the total heat input to the bottom stripper. This will include the heat into the top stripping tray plus the heat in the stripping steam (enters below the bottom tray).

Step 3. Calculate the heat out of the stripper. This includes the bottom product at an unknown temperature and enthalpy, the stripout from the top stripping tray, and the steam from the top stripping tray. Assume the stripout and the steam will be at 5°F below the flash zone temperature and the steam will be at the flash zone partial pressure for the steam.

Step 4. From the equation

Heat in = Heat out

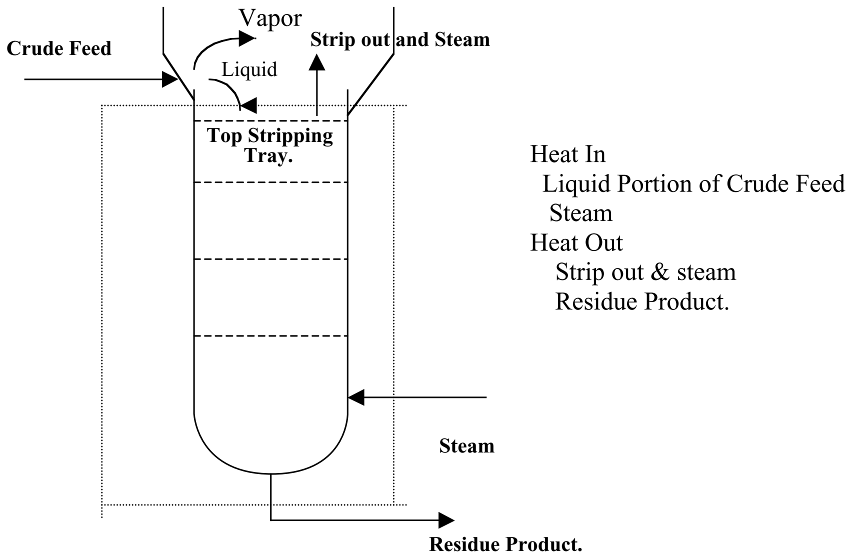


Figure 3.11. Heat balance diagram for below feed tray.

calculate the Enthalpy (in Btu/lb) and thus the temperature of the residue or bottom product leaving the tower.

Note: While the author uses enthalpy data from his own files, it is recommended that engineers use their own proven data from their own files, or data contained in Maxwell's *Hydrocarbon Data,* or those given in the *GPSA Engineering Data Book.*

With all the temperature and enthalpy criteria established an overall heat balance over the tower can proceed. In this balance the heat in with the crude will be that for the overflashed crude at the flash zone temperature. The other heat inputs are the steam streams to the various strippers. The heat out will consist of the overhead distillate product at the distillate drum temperature, the overhead condenser(s) duty, the condensed steam, all the side streams and residue as liquid and at their respective stripper bottom temperature, and finally the total of the pumparound duty.

As all the Enthalpy duties are known except that for the pumparound this can now be calculated from the expression

$$\text{Heat in} = \text{Heat out.}$$

Distributing the total pumparound to meet the required internal reflux

The degree of separation of one product from an adjacent one is determined by, the number of fractionating trays used for the separation and the liquid overflow (Reflux)

entering the section used for the separation. As has been said previously the degree of separation of products in the Atmospheric Crude Distillation tower is measured by the difference between the 95 %vol point of the lighter fraction and the 5 %vol point of the heavier fraction as measured by their ASTM distillation curve. Packie developed a relationship between the ASTM distillation difference, the number of trays, and the amount of overflow in terms of internal reflux ratio. This relationship is given by a series of curves in the appendix to this chapter (Figure 1.A.2). The use of these curves is illustrated by the following stepwise calculation procedure to arrive at the amount of overflow liquid.

Step 1. Establish the ASTM gap (or overlap) that is required. This is usually given in the design specification of the unit or by some constraint on one or other of the products. It is well to note here that ASTM gaps can usually be achieved between products lighter than light gas oil. Gaps can rarely be achieved within the tower for products heavier than the light gas oil cut. To trim the front or back ends of heavy products may require additional equipment such as vacuum flash equipment.

Step 2. Establish the number of trays separating the two adjacent products. This has usually been fixed by the design of the tower in operating units. For design purposes the following Fractionating number of trays can be used as a start point and guide:

	Tray number
Top tray	1
Kero draw off	12
LGO PA return	19
LGO draw off	22
HGO PA return	29
HGO draw off	32

As a rule of thumb the section of trays for the pumparound may be counted as one fractionating tray. Thus for example $10 + 1$ trays can be counted for the separation of Kero and LGO.

Step 3. Evaluate the 50 %vol temperature difference between the TBP of the two adjacent cuts.

Step 4. Using the side stream to side stream family of curves in Figure 3.A.2 for max steam read off the factor corresponding to the ASTM difference required on the appropriate $\Delta 50\%$ TBP temperature curve.

Step 5. The factor read is the number of trays separating the two products' draw-off multiplied by the Reflux ratio. This Reflux ratio is defined as:

$$\frac{\text{Hot overflow in gallons per hour leaving the lightest product draw-off tray}}{\text{Total product gallons per hour entering the lightest product draw-off tray}}.$$

Equate to determine the overflow quantity, in hot gallons per hour. The units material balance is used to resolve the flows to lbs per hour.

It now only remains to utilize the overflow criteria in the overall tower heat balance to arrive at a duty for the respective pumparound. If there is only one other pumparound in the tower it can be determined by subtraction of the calculated pumparound from the total PA duty. Should there be more than two pumparounds involved then the calculation steps given above must be repeated for the second one, and the third obtained by subtraction.

On completion of setting the individual pumparound duties as described the items determined by subtraction must be checked to ensure that they do in fact generate enough overflow to meet the fractionation criteria. If they do not fulfill this then the overall pump: around duty must be increased. This can be achieved only by increasing over flash if the cut points are to remain the same. Should the check calculations show there is somewhat more internal reflux than required, providing it is not excessive, it should remain as is. If it is excessive the over flash can be reduced accordingly.

Checking the number of trays allocated to the pumparound

With the individual pumparound duties now established, the temperatures to and from the pumparound section trays are the next items to be fixed. Use the following steps to accomplish this:

Step 1. Draw the tower temperature profile from the flash zone to the tower top tray. Use the already calculated side stream draw-off temperatures for this.

Step 2. Assume the vapor temperature to a tray is the liquid temperature on the tray below it. This is a reasonable assumption and well within the accuracy required.

Step 3. Fix the pumparound inlet temperature. This will need to be a guess at this point, and may well change on completion of the heat exchange system calculations. From the pumparound duty calculate the flow of pumparound liquid in lbs per hour.

Step 4. Carry out a heat balance over the pumparound section of the tower. This will now include the following items:

Heat in

With the HC vapor from the tray below the section (includes the overflow).
With steam from the tray below the section.
With steam and strip out from the product stripper (if applicable).
With the pumparound return liquid.

Heat out

With lighter product vapors rising from the top PA tray.
Total steam from flash zone and side stream strippers below top PA tray.
Liquid pumparound leaving the tower.
Liquid product draw off plus strip out leaving to stripper (if applicable).
Liquid overflow from the PA draw-off tray.
The pumparound duty in Btu/hr.

In these items the only unknown is the enthalpy and therefore the temperature of the vapor leaving the top pumparound tray. Equate Heat in = Heat out to solve the unknown.

Step 5. With the duty of the pumparound section in Btu/hr and the temperatures in and out now known, the pumparound section can be treated as a heat exchanger. The area of heat transfer will be the total tray areas of the section, and the heat transfer coefficient in terms of Btu/hr/sqft/°F is given in Figure 3.12.

Step 6. The area is now calculated from the heat transfer equation;

$$Q = UA \quad \Delta t_m$$

where

Q = Heat transferred in Btu/hr
U = The overall heat transfer coefficient in Btu/hr/sqft/°F
A = Heat transfer area in sqft.
Δt_m = Log mean temperature difference in °F.

Step 7. The number of trays required will be determined from the tray or tower diameter calculation provided by a future calculation to determine the tower dimensions.

Calculating the main tower dimensions

Having established the number of trays and the relative location of pumparound and side stream draw off, the overall dimensions of the tower can be calculated. Note: The number of trays allocated to the pumparound sections may be revised based on the tray or tower diameter calculation for those sections. This will not affect the tray loading calculations which will determine the tray diameters. The calculation to determine the tower diameter(s) is based on the vapor/liquid loading on each section of the tower. This loading is based on a series of heat balances to determine this traffic to and from the critical trays. These critical trays are:

- The top stripping tray of the residue stripping section (below the flash zone)
- The bottom side stream and pumparound draw-off tray
- The next pumparound draw-off (and the side stream if applicable) tray

Units are in BTU/h. °F. Ft^2 Fluid being cooled	Fluid being heated	U_o
Exchangers		
Naphtha pumparound	Crude	70–80
Kero	Crude	70–75
Debutanizer bottom	Debutanizer feed	75
Gas oil (inc. BPA)	Crude	40–50
Reduced crude	Crude	20–30
Light end bottoms	Light end feed	70–75
Vacuum distillates	Crude	30–40
Bitumen	Crude	20
Cat oil slurry	FCCU feed	40
CDU overheads	Crude	80–90
Coolers		
Debutanizer bottoms	CT water	75
Light naphtha	CT water	80
Gas oils	CT water	40
DEA or MEA	CT water	110
Reduced crude	CT water	30
Vapour heat exchangers		
Reformer effluent	Naphtha feed	38
Reformer effluent	Recycle gas	38
Crude tower overheads	Crude oil feed	50
Condensers		
CDU overheads	CT water	65
Splitter overheads	CT water	85
Amine stripper overheads	CT water	100
Debutanizer overheads	CT water	90
Reformer effluent	Water	65
Air coolers		
Naphtha coolers		23
Debutanizer bottoms		30
Light end overhead		40
Reboilers (use heat flux in BTU/h/ft^2)		
	Hydrocarbon	Water
18 in. dia. bundles	20 000	30 000
30 in. bundles	17 500	26 500
>30 in. bundles	15 000	23 000

Figure 3.12. Overall heat transfer coefficients in oil refining.

- Any other lighter pumparound draw-off tray
- Any lighter side stream draw-off tray
- The top tray

The loading below the bottom pumparound and product draw-off gives the vapor loading for setting the *maximum* tower cross sectional area and diameter. This is fixed by the crude feed cut point and the degree of over flash.

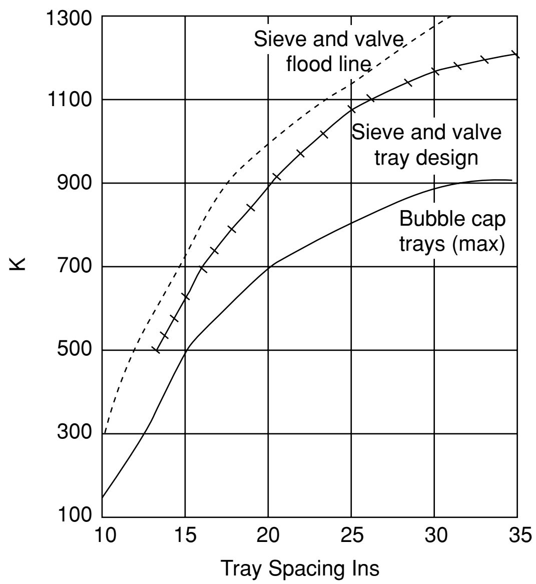


Figure 3.13. Values for K_f.

Calculating the tower vapor and liquid loading

These loadings at the critical trays are calculated from a series of heat balances starting at the bottom stripping section and continuing up the tower to the top fractionation tray. Consider the loading on the bottom side stream and pumparound draw-off tray. This commences with a heat balance from below this tray and includes the streams shown in the envelop in Figure 3.23.

The heat into and out of the envelope as shown is provided by the following streams:

Heat in: The crude vapor and liquid feed at the flash zone conditions.
The steam to the residue stripper.
The overflow liquid from the draw-off tray.
Heat out: The residue product.
The vapor of the total distillate products from the tray below the draw off.
The overflow vapor from the tray below.
The steam from the flash zone at the conditions of the tray below.

The temperature and quantities of these streams are known (from the material balance), except the quantity of overflow. Their temperatures are also known from the tower temperature profile or from previous calculations. To complete the calculation let the

lbs of overflow be x and give all the streams their enthalpy value. Then solve the equation:

$$\text{Heat in} = \text{Heat out}$$

to find the value of x.

Repeat the heat balance calculation for the envelop ending below the next pumparound draw-off tray. In this case the vapor leaving the envelop will contain the overflow vapor from the pumparound tray, and all the distillate products except the heavier product drawn off from the pumparound tray below. It will also contain all the steam from the flash zone *and* from the side stream stripper of the heavier product. The remaining heat out of the envelope will be the stripped heavier product and the bottom pumparound duty in Btu/hr. The heat into the envelope will be as before except this time the unknown overflow quantity will be that flowing from the lighter pumparound draw-off tray. It will also include the steam from the heavier product side stream stripper.

The tower diameter. The tower diameter may now be calculated from the vapor flow to the tray and the total liquid load on the tray. The quantity of the vapor to the tray has already been established. The liquid load on the tray will be:

The product draw-off liquid.
The pumparound liquid.
The calculated over flow liquid.

These quantities must be in terms of weight and volume at the tray conditions of temperature and pressure.

The calculation procedure and data described below will give a good *estimate* of the tower diameter. For design purposes however the tray fabricator/designer's data and procedure should be used. Trays are usually proprietary items covered by patents, their performance therefore are subject to guarantees.

A calculation procedure to estimate the tower diameter is as follows:

Step 1. Summarize the liquid traffic through the tray.
Step 2. Select the type of tray that is to be considered for the design. (in the case of an existing unit refer to the fabricators drawings). Valve and sieve trays have reasonable similarity in their major characteristics. Bubble cap trays are seldom used these days. A table of the valve tray characteristics is given in the appendix as A.1.3.
Step 3. Compute the liquid loading on the tray being checked for size. This will include *all* the liquid entering the tray. For example on a side stream draw-off tray

under a pumparound section it will include;

- The side stream feed to the stripper
- The pumparound liquid
- The liquid overflow from the tray

This loading should be in cubic feet per second (CFS).

Step 4. In the case of a new design, set the downcomer area in accordance with Table A.1.3. For an existing trays use fabricator's drawings. Calculate the linear velocity of the liquid in ft/sec. For good design this velocity should not exceed 0.6 ft/sec at the downcomer outlet. Tray spacing should be such that the liquid level in the downcomer should not exceed 50% of the tray spacing. To meet this criteria tray spaces on these critical liquid loading trays may be higher than the remaining tray spacing in the tower.

Step 5 . Summarize the vapor traffic to the critical tray under examination. This should be in lbs/hr and moles/hr. The vapor should be the total vapor as used to calculate the tray overflow.

Step 6. Calculate the flood vapor velocity Gf in lbs/hr sqft. For good tray design (or performance of an existing tray) the actual vapor velocity should not exceed 90% of this flood value. This vapor flood velocity is calculated using the following expression:

$$G_f = K_f \sqrt{\rho_v} \times (\rho_l - \rho_v)$$

where

G_f = Mass velocity in lbs/hr sqft of bubble area at flood.
ρ_v = Density of vapor at the tray conditions in lbs/cuft.
ρ_l = Density of liquid at the tray conditions in lbs/cuft.
K_f = constant based on tray spacing and given in Figure 3.13.

Step 7. Using the actual vapor load per sqft of bubble area as 90% of G_f, calculate the bubble area as G_a divided into the total vapor flow.

Step 8. Establish the following criteria using the characteristics given in Table A.1.3.

Where:

A_s = Total tray area in sqft.
A_{dc} = Inlet and outlet down comer areas in sqft.
A_w = Waste area (about 20% of A_b).
A_b = Bubble area.
$A_s = A_b + A_w + A_{dc}$

Use Figure 1.A.4 (relationship of chord height, area, and length) in the appendix as applicable.

Calculate the overall tower height

Using the calculated vapor and liquid calculations check and fix the number of trays used to meet the tower's fractionation requirements. Then check the tray spacing. For normal fractionation a 24 inch tray spacing is acceptable. For draw-off trays with pumparound the section should be checked for high downcomer filling (that is in excess of 50%). Trays in this section often require bigger tray spacing, usually 30 inch. Calculate the tower height allocated to the trays. Allow 6 ft between top tray and the top tangent line of the vessel to allow space for the reflux distributor and good liquid/vapor separation.

The space required for the flash zone (between the bottom fractionating tray and the top residue stripper tray) is usually 10–14 ft depending on throughput and the swirl size. The hot well below the bottom residue stripping tray should be sized to allow for the steam inlet distributor and a surge hold up of the residue product. This surge hold up is based on the company's operating policy, but as a guide:

- If the product is routed to storage 5 min hold up.
- If the product is to be fed into a fired heater of a downstream unit allow 15 min hold up.

The stripper column

This column is sized using the same procedure as used for the main column sizing. As mentioned earlier the strippers for the various side streams are usually stacked one above the other to form a single uniform diameter tower. The only diameter that may change is that for the hold up section in the bottom product stripper. This may be increased to adjust the height of the column to ensure the free flow of un stripped feed to each stripper from its respective draw-off tray in the main tower.

The number of trays for steam stripping is usually four. The tray spacing is also usually 24 inches. For the height of the tower begin with setting the height of the bottom stripper's bottom tangent line above grade. This should be at least 15 ft to allow a reasonable available NPSH for the product pump. From there allow a 5–15 min surge capacity at the calculated tower diameter and some 12 inches for the steam inlet distributor below its bottom stripping tray. Note the same comments apply to surge capacities in this tower as given for the residue stripper product.

The crude feed preheat exchanger system design

All crude distillation units pre heat the incoming crude oil feed by heat exchange with hot product and reflux streams. The preheated crude is then partially vaporised to satisfy the flash zone conditions by a fired heater. The degree of preheat is a question of an economic balance between the cost of the hardware and the savings in utilities

due to the recovery of heat from the hot product and reflux streams. The exercise to arrive at this economic balance is one of critical analysis of the respective enthalpy and its temperature levels of the various streams to be considered for heat exchange. It requires also good cost data for the heat equipment included in the system. This equipment must include the cost of the heat exchangers (usually shell and tube), cost of the final product coolers, and the cost of the final crude feed fired heater.

More than one scheme is developed and their capital costs calculated from up to date equipment vandor data if possible. Next the utility requirements and cost for each system is developed using the company's unit utility costs. For an examination and comparison of these systems these cost data must be on the same bassis for each of the schemes. The following procedure is one of several that can be used for this economic analysis:

Step 1. Construct the enthalpy curve for the crude feed. This stream will remain in the liquid phase through the preheat train. The enthalpy curve therefore needs only to consider sensible heat. Start at say 60°F, calculate the enthalpy by multiplying the Btu/lb (from the enthalpy curve) by the total weight in lbs/hr, this gives the enthalpy point on the curve in Btu's/hr. Proceed with a further 10–15 other temperature points to make a smooth temperature V's enthalpy curve.

Step 2. Examine the temperature and weight/hr of the overhead, pumparound, and product streams. Select those that will be candidates for heat exchange against the crude feed. For example, the overhead stream from the main tower is a prime candidate for heat transfer against the cold crude because of its high heat content (latent heat) over a wide range of temperature. The kerosene product stream on the other hand is usually a poor candidate. It contains only sensible heat over a short temperature range and is also probably the smallest product stream by weight.

Step 3. Prepare enthalpy curves for the selected candidates. In the case of pumparounds and the side stream products only the enthalpy (in Btu/hr) against say three or four temperatures need to be plotted.

Step 4. The overhead vapor enthalpy curve requires a more complex calculation. Here the enthalpy curve must be based on its condensation curve. This requires a calculation of the streams equilibrium (vapor and liquid) composition at the condensing range of temperatures and pressures. For this purpose assume a straight line pressure drop and temperature profiles between tower top and reflux drum. Select four to five pressures and their corresponding temperatures. For the purpose of this calculation it is assumed that no steam condenses in this segment of the system, therefore the equilibrium constants are taken at the partial pressure with steam. Using the overhead vapor composition (in moles/hr) from the sum of overhead product plus reflux vapor calculate its vapor/liquid composition at the selected temperatures and pressure. Apply the enthalpy value to both phases (not forgetting the steam) for the selected temperatures. Plot enthalpy in Btu/hr versus the selected temperature.

Step 5. Superimpose these product, overhead, pumparound enthalpy curves on the crude feed enthalpy curve. Start with the overhead curve and then with each other stream in their process sequence. Draw these enthalpy curves to provide a reasonable temperature approach to the crude feed curve. Figure 3.14 is an example of this concept.

Reasonable temperature approaches should not be less than 20°F for distillates and between 40 and 60°F for residues and heavy distillates.

Step 6. Several different schemes can now be developed using a ruler and set square for each of the product stream in different sequences. In the case of the exchange of heat against large volume streams such as the residue and perhaps the lower pumparound these streams can be split. They can also be shown to flow in series against the crude or in parallel.

Step 7. Size the heat exchanger equipment required for each of the schemes developed. This includes the sizing for additional equipment for each stream to meet its required end temperature. For example the final air coolers to meet product rundown temperature or trim coolers to meet pumparound return temperature. Sizing these items need not be precise as long as they are on the same basis. The simple equation for heat transfer

$$Q = UA\Delta t_{\mathrm{m}}$$

will suffice for this purpose. The value of U may be taken from Figure 3.12.

Step 8. Note the end temperature of the crude and using this enthalpy and the total enthalpy in the crude at the flash zone of the main tower determine the duty required by the fired heater in each case.

Step 9. Cost out the equipment required for each scheme. This includes all the heat exchangers (including trim and final coolers), and the fired heater. These can usually be obtained from equipment vendors on a unit size ($ per sqft for example) basis. Estimate the utility requirements for the equipment in each scheme.

Step 10. Tabulate the results starting with the highest capital cost scheme as a base case and calculate a simple incremental return on investment based on savings for the remaining schemes. The scheme with the highest positive ROI should be the system selected.

In almost all modern crude distillation units there are facilities to remove free salt. These desalting facilities are proprietary units which consist of fresh water injection into the crude feed and subsequent separation of the water, with the salt now in solution, from the oil. This separation takes place at a fixed temperature at the appropriate point in the crude feed preheat train. Demulsifying chemicals or (more usually) electrostatic precipitators or both are used to enhance this separation process. The temperature for this desalting process (usually around 220–250°F) must be catered for in the crude feed preheating.

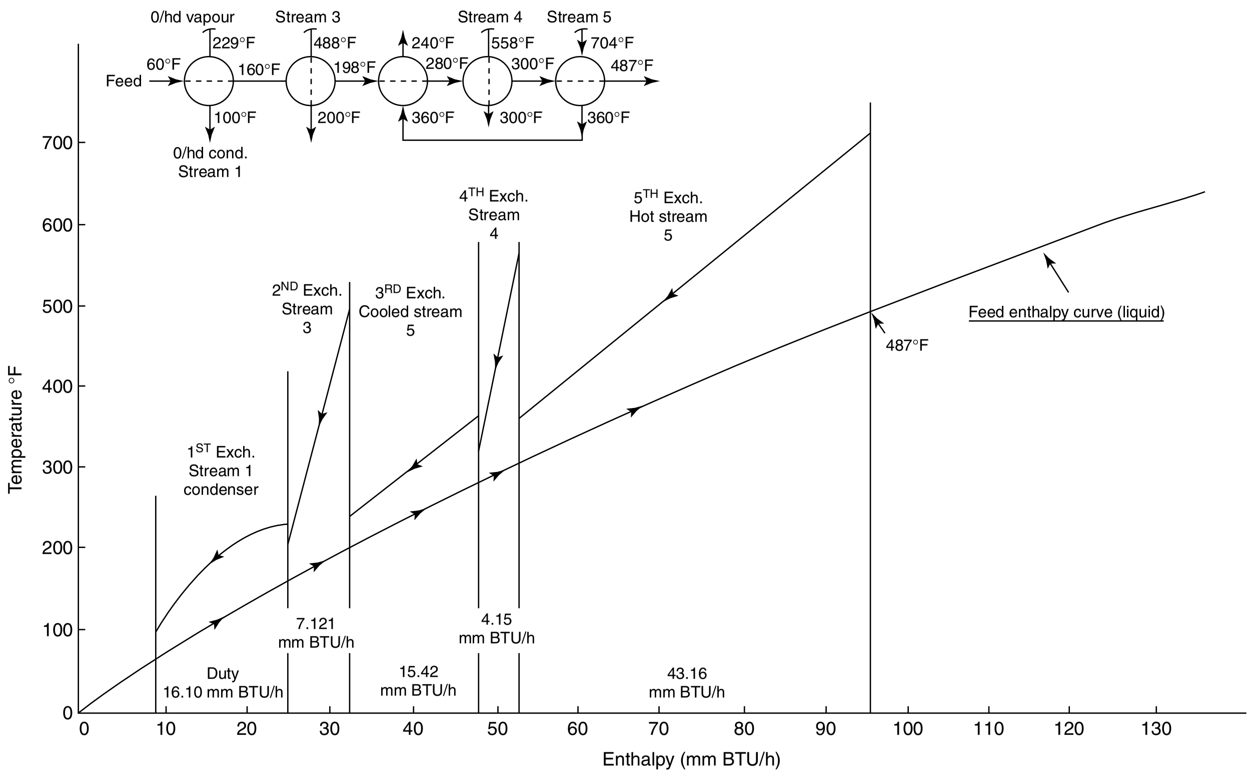


Figure 3.14. Calculation of the crude preheat train.

An example in the design of an atmospheric crude oil distillation tower

This example is confined to the design of the main fractionator and the associated stripper column. It is based on the processing of 30,000 BPSD of Kuwait crude providing the following products:

Overhead full range naphtha	Gas to 375°F
1st side stream kerosene	375–480°F
2nd side stream light gas oil	480–610°F
3rd side stream heavy gas oil	610–680°F

The distillation specifications shall be as follows:

- The ASTM end point of the Naphtha shall not exceed 400°F.
- The difference between the ASTM 95% point of the Naphtha and the 5% point of the Kero shall be at least 25°F.

The difference between the ASTM 95% point of the kerosene and the 5% point of the Light Gas Oil shall be −10°F (10° overlap). The ASTM end point of the Light Gas Oil shall not be greater than 620°F. The difference between the ASTM 95% point of the Light Gas Oil (LGO) and the 5% point of the Heavy Gas Oil (HGO) shall not exceed −35°F (35° overlap). The FBP of the TBP shall not exceed 710°F and a Condrason Carbon content of 8.0% maximum.

Developing the material balance

The crude TBP and sg curves are given in Figures 3.15 and 3.16, respectively. The TBP curve is divided into the distillate product cuts and also into narrow boiling point components.

The ASTM curves for the cuts are developed from the product cut points and probability paper. These are converted to TBP using the 'Edmister' method as shown in Table 3.1.

These values are plotted on Figure 3.17 and the narrow range components indicated for each of the product cuts. Note the front end of the naphtha is the front end of the whole crude so only the last segment of the cut is required in this calculation.

The cut characteristics are shown in Table 3.2.

Overflash Cut range 690–725 (3 %vol on crude)

Sg @ 60 °F	0.891
Mol wt	295.

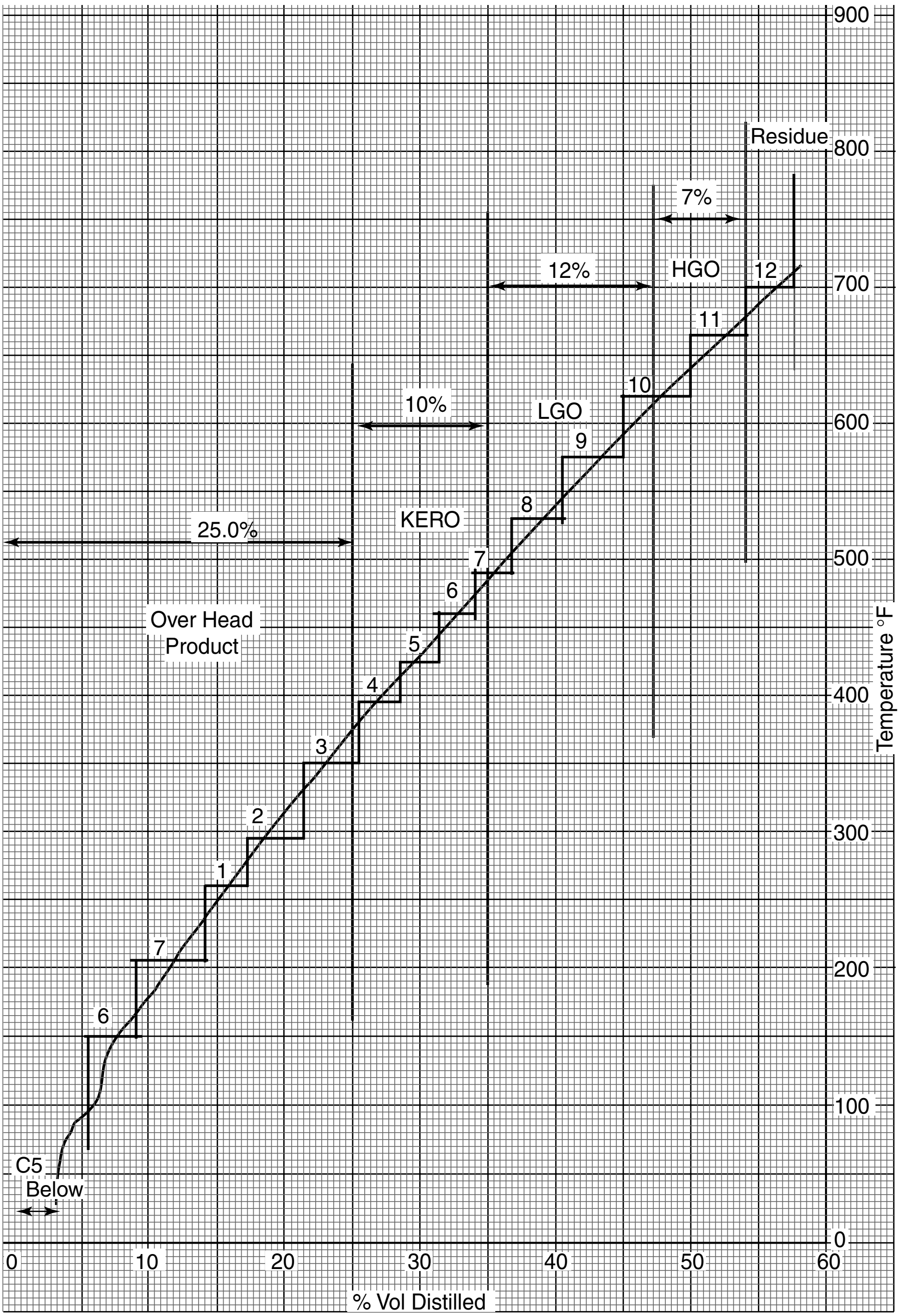


Figure 3.15. The TBP curve for the example calculation.

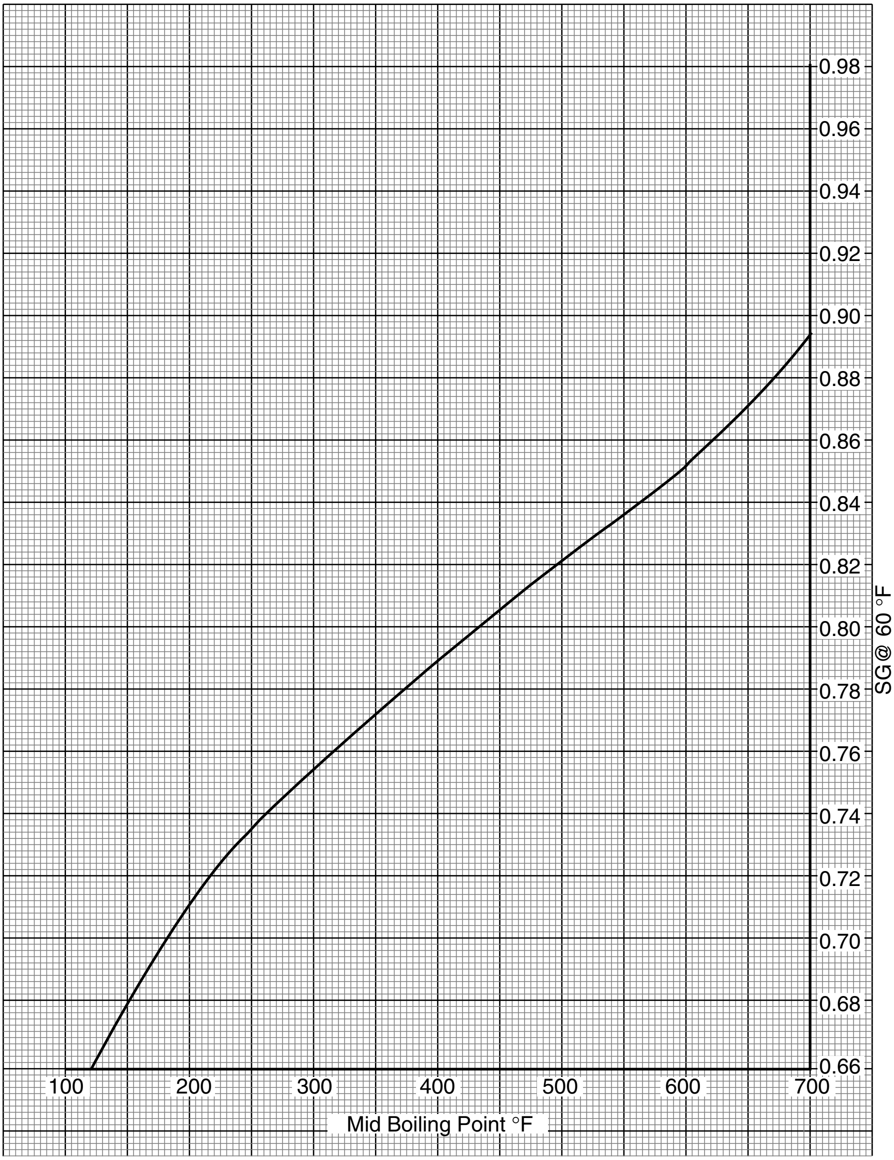


Figure 3.16. The SG curve for the example calculation.

Table 3.1.

	ASTM, °F	Δt ASTM, °F	Δt TBP, °F	TBP, °F
Naphtha (overhead)				
IBP		Not Required		
10%				
30%				
50%	300	10	14	305
70%	310	20	27	319
90%	330	28	32	346
FBP	358			378
Kero (1st side stream)				
IBP	345	31	57	287
10%		124	44	344
30%	400	19	32	388
50%	419	17	27	420
70%	436	24	33	447
90%	460	35	38	480
FBP	495			518
LGO (2nd side stream)				
IBP	495	25	48	457
10%	520	12	25	505
30%	532	11	20	530
50%	543	16	26	550
70%	559	13	19	576
90%	572	23	27	595
FBP	595			622
HGO (3rd side stream)				
IBP	595	20	40	568
10%	615	11	25	608
30%	626	7	14	633
50%	633	7	12	647
70%	640	18	25	659
90%	658	17	20	684
FBP	675			704

The complete material balance can now be written as shown in Tables 3.3.

Flash zone calculations

Total pressure at the flash zone
Estimate the overhead reflux drum pressure as 5 psig (This will be checked later by a bubble point calculation at 100°F).

Give the overhead air condenser a pressure drop of 5 psi. (This is a reasonable pressure drop for this equipment, and will be specified as such in the equipment data sheet to vendors.) Give the crude to overhead vapors heat exchanger a pressure drop of 2 psi (overhead vapors flow shell side).

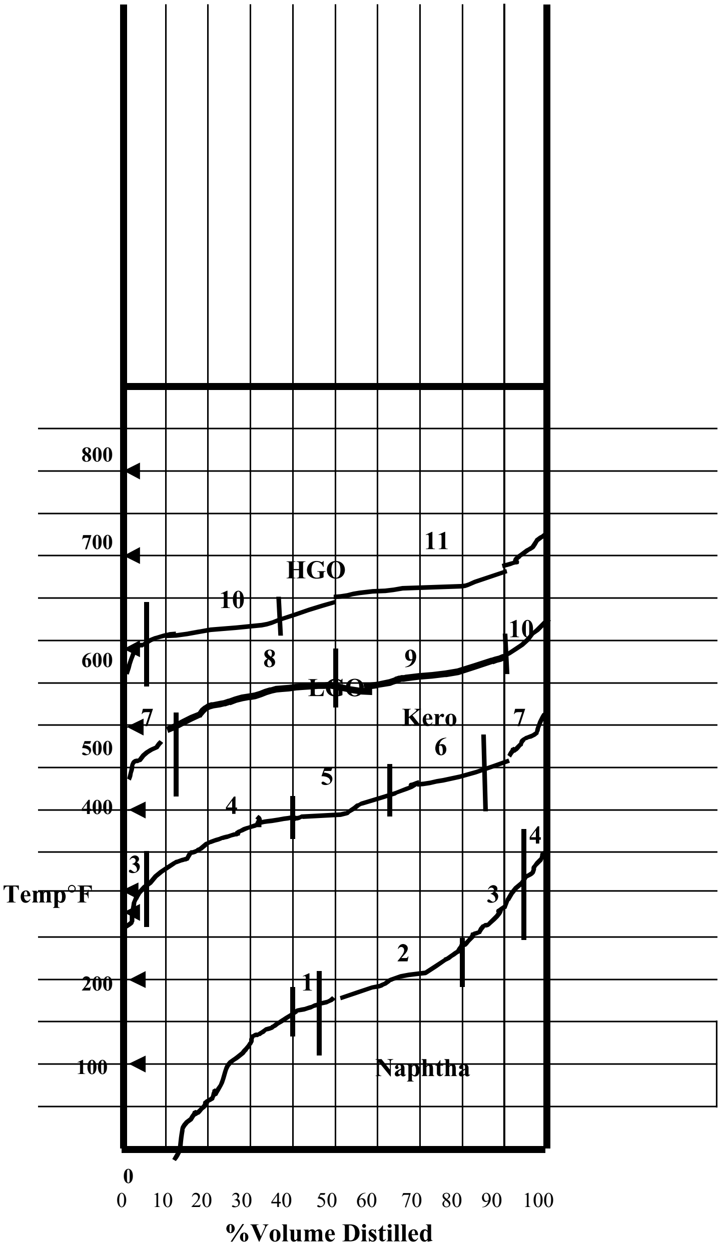


Figure 3.17. Product curves and narrow range components.

Table 3.2. Cut characteristics

Component	%vol on crude	SG @ 60°F	Wt factor	Mol wt	Mol factor
Overheads cut from gas to 375°F TBP cut point (25 vol% on crude)					
C2	0.11	0.374	0.411	30	0.014
C3	0.84	0.508	4.267	44	0.1
Ic4	0.40	0.563	2.252	58	0.039
nC4	1.53	0.584	8.935	58	0.154
C5's	3.02	0.629	18.996	72	0.264
C6	3.60	0.675	24.300	86	0.286
C7	4.50	0.721	32.445	100	0.324
Comp 1	3.50	0.743	26.005	114	0.228
2	3.77	0.765	28.841	126	0.229
3	2.41	0.776	18.702	136	0.138
4	1.32	0.788	10.402	152	0.068
TOTAL	25.0	0.702	175.556	95.2	1.884
Kerosene cut range 375–480°F (10 %vol on crude)					
2	4.0	0.765	30.6	126	0.24
3	14.0	0.776	108.64	136	0.80
4	22.0	0.788	173.36	152	1.14
5	24.0	0.799	191.76	165	1.16
6	21.0	0.810	170.10	177	0.96
7	13.0	0.825	107.25	190	0.56
8	2.0	0.839	16.78	205	0.08
TOTAL	100.0	0.798	798.49	161.6	4.94
Light gas oil cut range 480–610°F (12 %vol on crude)					
6	2.0	0.810	16.2	177	0.09
7	9.0	0.825	74.25	190	0.39
8	38.5	0.839	323.02	205	1.58
9	40.5	0.848	343.44	238	1.51
10	10.0	0.860	86.00	250	0.34
TOTAL	100.0	0.843	842.91	215.6	3.91
Heavy gas oil cut range 610–680°F (7.0 %vol on crude)					
9	6.0	0.848	50.88	228	0.22
10	33.0	0.860	283.80	250	1.14
11	61.0	0.887	540.77	287	1.88
TOTAL	100.0	0.875	875.45	270.2	3.24

Neglect the pressure drop for fittings and piping—this will be quite small for a properly designed unit.

Then total tower top pressure is 12 psig.

Assume 40 valve trays in the section of the tower between flash zone and tower top. Let the pressure drop per tray be 0.25 psi. Then pressure drop in this section of the tower is 10 psi.

Table 3.3. The material balance

Stream	Cut range	%vol	Cum %vol	BPSD	GPH	SG	#/Gal	lbs/hr	%wt	Mol wt	Mol/hr
Whole crude	–	100.0	100.0	30,000	52,500	0.8685	7.23	379,575	100.0	225.3	1,684.8
O/head	IBP −360	25.0	25.0	7,500	13,125	0.702	5.84	76,650	20.2	95.2	805.1
Kero	−480	10.0	35.0	3,000	5,250	0.798	6.64	34,860	9.2	161.6	215.7
LGO	−610	12.0	47.0	3,600	6,300	0.843	7.02	44,226	11.7	215.6	205.1
HGO	−690	7.0	54.0	2,100	3,675	0.875	7.28	26,754	7.0	270.2	99.0
Resid	+690	46.0	100.0	13,800	24,150	0.957	8.16	197,085	51.9	547.6	359.9
The flash zone material balance											
O/flash	−725	3.0	57.0	900	1,575	0.891	7.4	11,655	3.0	295	39.5
Prod vapor	−690	54.0	54.0	16,200	28,350	0.773	6.43	182,490	48.1	137.7	1,324.9
Total vap	−725	57.0	57.0	17,100	29,925	0.780	6.49	194,145	51.1	142.3	1,364.4
Resid*	+725	43.0	43.0	12,900	22,575	0.988	8.22	185,430	48.9	578.7	320.4
Total		100.0	100.0	30,000	52,500	0.8685	7.23	379,575	100.0	225.3	1,684.8

*Does not include liquid overflow from bottom wash tray.

Total flash zone pressure is 12 psig + 10 psi = 22 psig. Call it 25 psig (40 psia) for design purposes.

Calculate the partial pressure of the hydrocarbon vapor at the flash zone
Take the quantity of stripping steam as 1.2 lbs/gal of residue (from Figure 3.10).

$$\text{The lbs/hr of stripping steam is } 1.2 \times 24{,}150 = 28{,}980 \text{ lbs/hr.}$$
$$= 1{,}610 \text{ moles/hr}$$

The partial pressure of the hydrocarbon vapor therefore is:

$$\frac{\text{Moles HC vapor}}{\text{Total moles vapor}} \times \text{Total pressure} = \frac{1{,}364.4}{1{,}364.4 + 1{,}610} \times 40 \text{ psia} = 18.35 \text{ psia.}$$

Calculate the EFV curve of whole crude at atmospheric pressure
From the crude TBP curve, the slope of the whole curve is 11.8°F/%vol (10–70 %vol on TBP temperatures divided by 60). From the Maxwell curves the slope of the flash reference line slope is 8.5°F/%vol. (See Chapter 1 of this Handbook.)

$$\Delta T_{50\%}\,(\text{DRL} - \text{FRL}) = 40°\text{F}$$

$$T_{50\%}\,\text{DRL} = 667°\text{F}$$

$$\text{Then } T_{50\%}\,\text{FRL} = 667 - 40 = 627°\text{F}$$

Table 3.4. Flash curve at atmospheric pressure

	TBP				EFV		
%vol	Curve, °F	DRL	ΔT	Ratio	ΔT_1	FRL	Flash, °F
0	−127	75	−202	0.24	−48	200	152
10	190	190	0	0.4	0	285	285
30	420	430	−10	0.34	−3	450	447
50	645	667	−22	0.34	−8	627	619
70	900	900	0	0.34	0	795	795
90	1,235	1,140	95	0.34	32	915	947
100	2,192	1,250	942	o.34	320	1,040	1,360

Table 3.4 defines the flash curve at atmospheric pressure.

The flash curve calculated above is that at atmospheric pressure. To plot this at any other pressure take the 50 %vol temperature, and using the vapor pressure curves for hydrocarbons (see Figure 1.A.1 in the appendix) read off the temperature at the desired pressure. Draw the EFV curve through this new temperature parallel to the atmospheric curve. The flash zone temperature is the temperature at the % distilled on the partial pressure curve.

$$\text{Flash zone temperature} = 720°\text{F}$$

The EFV curve is shown in Figure 3.18.

Calculate total heat in the crude at the flash zone conditions
See Table 3.5 for the calculation of total heat in the crude at the flash zone conditions.

Please note: The enthalpy data used here are taken from the author's private files. It is recommended that the data given in Maxwell's *Hydrocarbon Data* or *The GPSA Engineering Data Book* be used for these heat balance calculations.

The tower heat balances

To calculate the temperature of the residue product leaving the tower
Consider the heat balance over the residue stripper as shown by the envelope in Figure 3.1.11. The product residue is 197,085 lbs/hr (from the material balance). The strip out vapor from the top stripping tray is 6% (from Figure 3.10) = 1,449 gals/hr. Assume the SG of the strip out is 7.5 lbs/gal (about the same as the overflash) and the mol wt is 305. Then the heat balance can be written as follows (see also Table 3.6):

Heat in = Heat out

$$197{,}085x + 45{,}404{,}000 = 119{,}916{,}000$$

$$x = 378 \text{ Btu/lb}$$

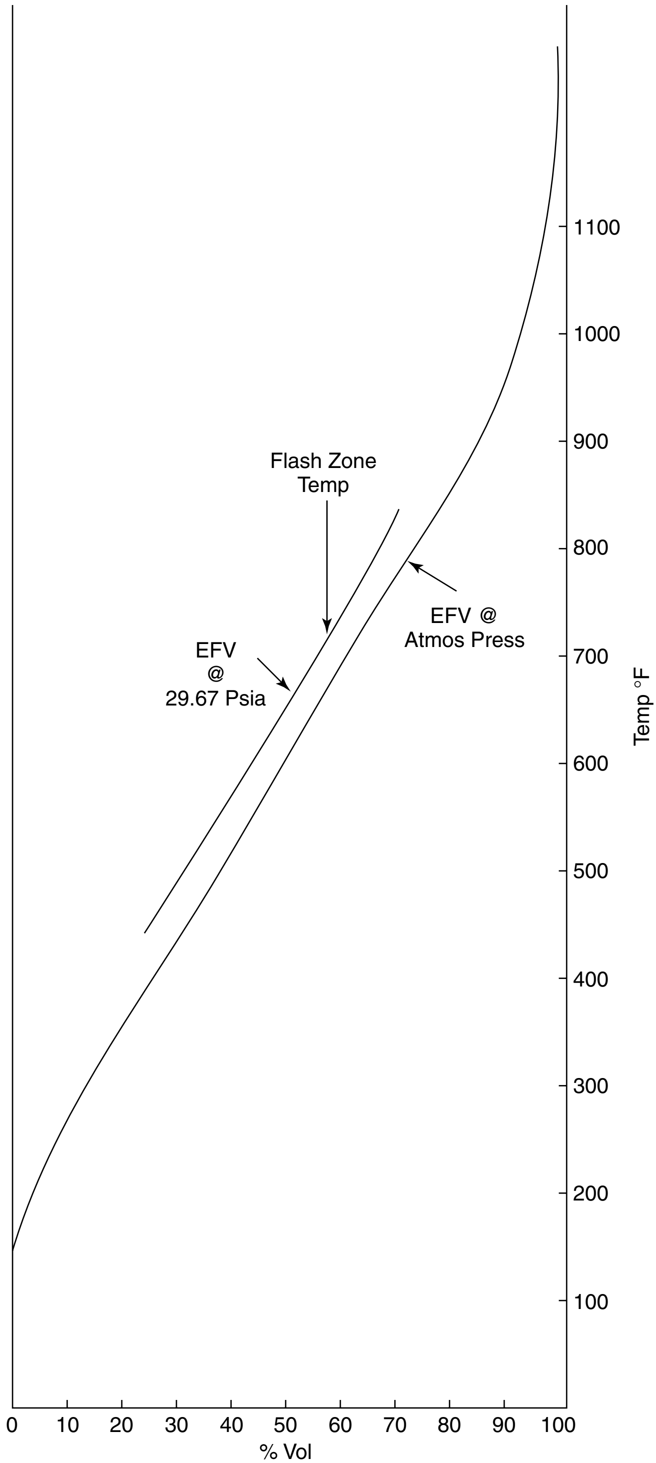


Figure 3.18. EFV curve for whole crude at flash zone conditions.

Table 3.5. Calculation of total heat in the crude at the flash zone conditions

Stream	V/L	°API	K	°F	Weight, lbs/hr	Btu/lb	mmBtu/hr
Crude vapor	V	5.0	11.8	720	194,145	528	102.509
Crude liquid	L	11.5	11.5	720	185,430	396	73.43
Total					379,575		175.939

From enthalpy tables this equates to 704°F.

To calculate the side stream draw-off temperatures
The steam rates to be used for side stream stripping will be:

All strippers will have three theoretical trays.
Both gas oil streams will use 0.5 lbs/gal respectively.
The Kero stripper will use 0.65 lbs/gal of steam.

Steam used is as follows:

Heavy gas oil $= 3{,}675 \times 0.5 = 1{,}838$ lbs/hr
Light gas oil $= 6{,}300 \times 0.5 = 3{,}150$ lbs/hr
Kero $= 5{,}250 \times 0.65 = 3{,}413$ lbs/hr

- Tower pressure profile.
  There will be 40 fractionating trays in the main tower (trays above the flash zone) and four residue stripping trays (trays below the flash zone). These trays will be numbered from the bottom to the top. Thus, the bottom residue stripping tray will be tray 1. The tower top tray will be tray 45. The pressure profile is shown in Figure 3.1.19.

Table 3.6.

						Enthalpy	
Stream	V/L	°API	K	Temp, °F	Weight, lbs/hr	Btu/lb	mmBtu/hr
IN							
Residue	L	13	11.5	720	197,085	396	78.046
Stripout	L	25.5	11.5	720	10,868	410	4.456
Steam	V			450	28,980	1,290	37.384
Total in					236,923		119.916
OUT							
Residue	L	13	11.5	t	197,085	x	$197{,}085x$
Stripout	V	25.5	11.5	715	10,868	490	5.325
Steam	V			715	28,980	1,383*	40.079
Total out					236,933		$45.404 + 197{,}085x$

*At partial pressure of the flash zone = 13.53 psig.

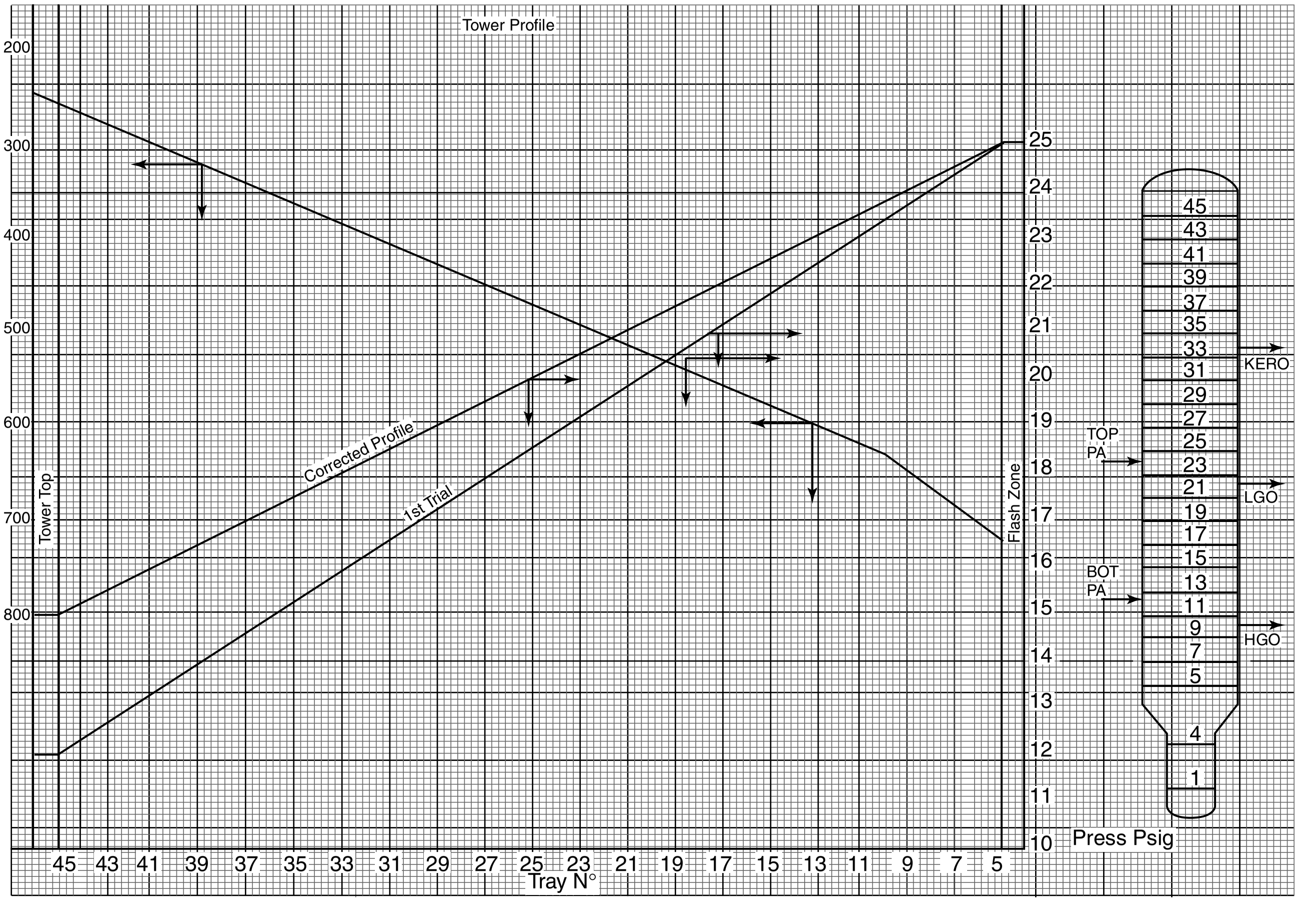


Figure 3.19. Tower pressure profile.

- Establish draw-off tray location
  Allow six wash trays above the flash zone to the HGO draw off. The HGO and bottom pumparound will be drawn from tray 10.
  Allow two trays for the pumparound return. Thus the pumparound will return on to tray 12.
  Allow 10 trays from tray 12 to the LGO and the top pumparound draw off. The draw-off tray will then be tray 22.
  Again allow two trays for the pumparound return. This will be tray 24.
  Allow 10 trays between top pumparound return and the Kero draw off. This locates the Kero draw-off at tray 34. It leaves 12 trays between the Kero draw off and the top tray (includes the draw-off tray and the top tray).
- Summary of tray locations and total pressure (1st trial).

	Tray no.	Pressure (psig.)
HGO draw off	10	23.4
BPA return	12	22.7
LGO draw off	22	19.5
TPA return	24	18.8
Kero draw off	34	15.5

- Calculate theoretical initial boiling points
  The FRL for each cut is developed from the TBP cut. Thus:

HGO	50% TBP is 648°F. Slope of TBP is 0.6°F/%vol. Slope of FRL is 0.2°F/%. $\Delta t50$ (DRL – FRL) is 7°F. 50% FRL is 641°F. From FRL curve IBP is 626°F.
LGO	This is developed in the same manner as HGO and the IBP of the FRL in this case is 527°F.
Kero	In the same way the Kero FRL has an IBP of 395°F.

  Now all these temperatures are at atmospheric pressure. It is now required to relate these temperatures to the partial pressure of the respective trays.
- To calculate the approximate partial pressures and draw-off temperatures

Assumptions:

1.0 All vapor lighter than the product cut is considered inert.

2.0 Internal reflux is assumed as follows:

To HGO tray 290 mol/hr.
To LGO tray 250 mol/hr
To Kero tray 200 mol/hr

$$\text{HGO tray partial pressure} = \frac{\text{Moles of HGO}}{\text{Total moles HC vapor + Steam}} \times \text{Total pressure}$$

$$= \frac{1,614.9}{3,224.9} \times 38.1 = 19.6 \text{ psia}$$

From vapor pressure curves theoretical draw-off temperature is 720°F.

From Figure 3.9:

Theoretical temperature − Actual temperature = 93°F
Actual draw-off temperature is 720 − 93 = 627°F

Draw-off temperatures for the LGO and Kero are calculated in the same way and are:

LGO = 493°F
Kero = 364°F.

To calculate the tower top temperature
Set the reflux drum temperature and pressure. In this case these will be set at 10 psig and 100°F. Taking the pressure drop across the exchangers and piping the tower top pressure will be 15 psig.

Fix the cold external reflux at 0.8 times the total moles overhead product;

The total moles HC in the overhead vapor is 1.8 × 805.1 = 1,449.18 moles/hr

Total moles steam in the overhead vapor is 2,076

Partial pressure of the hydrocarbons in the overhead vapor is:

$$\frac{1{,}449.19 \times 29.7}{3{,}525.18} = 12.2 \text{ psia}$$

The tower overhead temperature is the dew point of the hydrocarbons at the partial pressure and is shown in Table 3.7.

$K_2 = K_1 \times \mathrm{S}x = 0.135 \times 0.984 = 0.133 = 246°\mathrm{F}$ which will be the tower top temperature.

To calculate side stream stripper bottom temperatures
These are calculated by heat balances over the respective side stream strippers. The following criteria are used in these calculations:

Table 3.7.

Comp	Mol frac	1st trial K	@ 250°F $X = y/K$	Mol wt	Weight factor	SG @ 60	Volume factor
C2	0.008	Neg	Nil				
C3	0.054	73.77	0.001	44	0.044	0.508	0.009
iC4	0.021	35.24	0.001	58	0.058	0.563	0.010
nC4	0.084	27.05	0.003	58	0.174	0.584	0.030
C5	0.143	11.80	0.012	72	0.864	0.629	0.137
C6	0.155	4.590	0.034	85	2.89	0.675	0.428
C7	0.175	2.213	0.079	100	7.90	0.721	1.100
Comp 1	0.124	1.066	0.116	114	13.224	0.743	1.780
Comp 2	0.124	0.557	0.223	126	28.098	0.765	3.673
Comp 3	0.075	0.311	0.241	136	32.776	0.776	4.224
Comp 4	0.037	0.135	0.274	152	41.648	0.788	5.285
Total	1.000		0.984	129.7	127.676	0.766	16.675

	HGO	LGO	KERO
Total strip out % vol	5	8	8
Mol wt of strip out	230	180	120
SG of strip out	0.865	0.820	0.750

Only the heat balance calculation of the HGO stripper is given below (see also Table 3.8). The other two stripper calculations will be similar in form (Figures 3.20–3.22).

$$\text{Solving for x: } 26{,}754x = 8{,}997{,}000$$

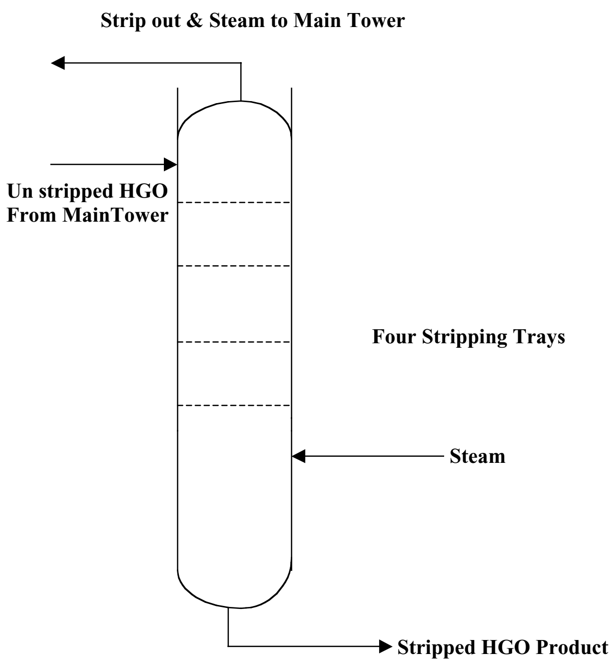
$$x = 336 \text{ Btu/lb}$$

From enthalpy tables = 615°F

Table 3.8.

Stream	V/L	°API	K	°F	lbs/hr	Btu/lb	mmBtu/hr
In							
Feed ex Hgo	L	30	11.5	627	26,754	347	9.284
Steam	V	–	–	450	1,838	1,290	2.371
Strip out	L	32	11.5	627	1,390	349	0.485
Total In					29,982		12.14
Out							
Hgo	L	30	11.5	t °F	26,754	x	$26{,}754x$
Steam	V			622	1,838	1,376*	2.529
Strip out	V	32	11.5	622	1,390	442	0.614
Total out					29,982		$3.143 + 26{,}754x$

*At partial pressure of 36.3 psia.



Heat In:
HGO Tray Draw off (Un Stripped HGO)
Steam

Heat Out:-
HGO Product
Strip Out
Steam.

Figure 3.20. Heat balance diagram over HGO stripper.

Temperature of LGO leaving its stripper = 488°F.
Temperature of Kero leaving its stripper = 359°F.

Overall tower heat balance

Tower overhead heat balances
Balance included by envelop 1 determines the overhead condenser duty. See Table 3.1.9.

The overhead condenser is 71.360, therefore the heat removed by intermediate refluxes (pumparound) is 115.717 − 71.360 = 44.357 mmBtu/hr.

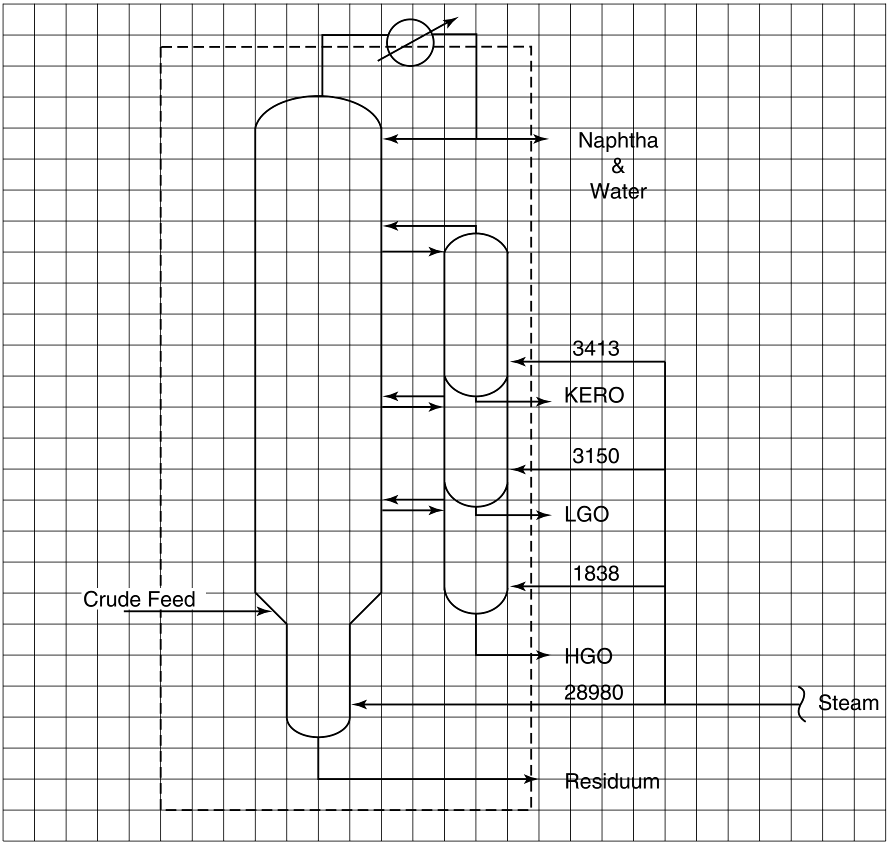


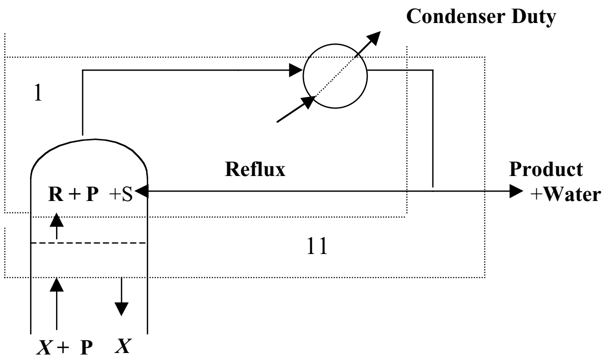
Figure 3.21. Overall tower heat balance diagram.

The heat balance included in envelope 2 determines the internal reflux from the top tray.

This is as follows:

Let x be the lbs/hr of overflow from the top tray. Then the heat balance is as follows (see also Table 3.10):

$$\text{Solving for } x = \frac{12{,}879{,}000}{121}$$
$$= 106{,}438 \text{ lbs/hr}$$
$$= 821 \text{ moles/hr or } 16{,}736 \text{ GPH.}$$



Envelope 1:-

Heat In

Reflux + Product Vapor

Steam

Heat Out

Reflux Liquid

Product Liquid

Water

Condenser Duty

Envelope 11:-

Heat In

Internal Reflux *X* Vapor

Product Vapor

Steam

Heat Out

Internal Reflux *X* Liquid

Product Liquid

Water

Condenser Duty.

Figure 3.22. Tower overhead heat balance diagram.

Check for fractionation: (using Figure 1.A.2 in the appendix)

Overflow @ 60°F = 16,736 GPH
Overflow @ 255° = 19,029 GPH
Prod vapor to top tray = 13,125 GPH
Reflux ratio = 1.45
Number of trays = 12
Factor is 12 × 1.45 = 17.4

50% TBP difference between Naphtha and Kero is 420 − 265 = 155°F.

ASTM Gap is 30°F which is within Spec.

Table 3.9a.

Stream	V/L	°API	K	°F	lbs/hr	Btu/lb	mmBtu/hr
IN							
Crude	V + L	–	11.7	720	379,575		175.939
Steam	V				37,381	1,290	48.221
Total in					416,956		224.160
OUT							
Residue	L	13	11.5	704	197,085	378	74.498
HGO	L	30	11.5	615	26,754	336	8.989
LGO	L	36.3	11.5	488	44,226	255	11.278
Kero	L	45.5	12	359	34,860	184	6.414
Naphtha	L	70.0	12	100	37,381	100	3.738
Refluxes							115.717
Total out					416,956		224.160

Table 3.9b.

Stream	V/L	°API	K	°F	lbs/hr	Btu/lb	mmBtu/hr
IN							
Naphtha	V	70	12	246	76,650	266	20.389
Reflux	V	70	12	246	61,230	266	16.311
Steam	V			246	37,381	1,197	44.745
Total in					175,351		81.445
OUT							
Naphtha	L	70	12	100	76,650	46	3.526
Reflux	L	70	12	100	61,320	46	2.821
Water	L			100	37,381	100	3.738
Cond duty							71.360
Total Out					175,351		81.445

Table 3.10.

Stream	V/L	°API	K	°F	lbs/hr	Btu/lb	mmBtu/hr
IN							
Naphtha	V	70	12	262	76,650	273	20.925
O/Flow	V	53	12	262	x	249	$249x$
Steam	V			262	37,381	1,199	44.82
Total In					$114,031 + x$		$65.745 + 249x$
OUT							
Naphtha	L	70	12	100	76,650	46	3.526
O/Flow	L	53	12	255	x	128	$128x$
Water	L			100	37,381	100	3.738
Cond duty							71.360
Total Out					$114.031 + x$		$78.624 + 128x$

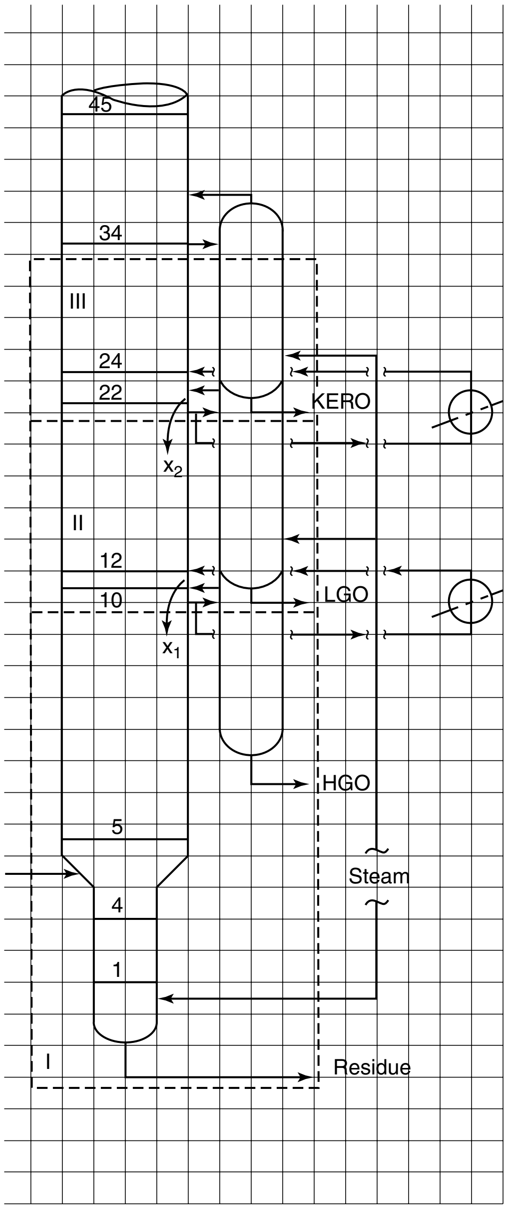


Figure 3.23. Remaining heat balance diagram for tower loading.

Calculating the remaining tower loadings
The tower loadings at the remaining critical sections of the tower are provided by heat balances over the envelopes given in the heat balance diagram (Figure 3.23). These critical sections are:

- Below bottom pumparound draw off (envelop 1)
- Below top pumparound draw off (envelop 2)
- Below Kero draw off (envelop 3)

Only the heat balance over envelop 3 is given in Table 3.11. The others follow in a similar fashion but the summaries of the calculated loads are given for all three sections (see Table 3.12).

$$\text{Solving for } x_3 = \frac{2{,}391{,}000}{111} = 21{,}540 \text{ lbs/hr.}$$

Checking fractionation:

Between HGO and LGO

$$\text{Reflux ratio} = \frac{22{,}809}{24{,}675} = 0.92$$

Number of trays = 11 (2 PA trays = 1 Frac tray).

Factor = 10.2

$\Delta 50\%$ TBP = 300°F

ASTM gap is +31°F which is well within the requirement for these two cuts.

Between LGO and Kero

Reflux ratio = 0.21
Number of trays = 11 (includes credit for PA trays.)
Factor = 2.3
$\Delta 50\%$TBP = 185°F
ASTM gap = −10°F

This is poor but as the kero cut is to be used as a blend stock it will be accepted . In the final design, however, provision will be made to increase the overhead reflux at the expense of the bottom pumparound to improve this separation.

Table 3.11. Heat balance over envelop 3

Stream	V/L	°API	K	°F	lbs/hr	Btu/lb	mmBtu/hr
ENV 3							
IN							
Heat in crude	V+L	–			379,575		175.939
Steam	V			450	33,968	1,290	43.819
Tray34 o/flow	L	46.3	12	367	$x3$	191	$191x3$
Total in					$413{,}543 + x3$		$219{,}758 + x3$
OUT							
O/flow to tray34	V	46.3	12	370	$x3$	302	$302x3$
Vapor to tray 34	V	62.5	12	370	111,510	320	35.683
Steam to tray 34	V			370	33,968	1,253*	42.562
LGO prod	L	36.3	11.5	488	44,226	255	11.278
HGO prod	L	30	11.5	615	26,754	336	8.989
Residue	L	13	11.5	704	197,085	378	74.498
Top PA							26.614
Bot PA							17.743
Total out					$413{,}543 + x3$		$217.367 + 302x3$

*Steam at the partial pressure of 18.3 psia.

Table 3.12. Summary of the main tower loading

	Liquid – From tray			Vapor – To Tray		
Draw-off tray	Lbs/hr	GPH	Hot GPH	Lbs/hr	Moles wt	Moles/hr
Tray 10 *		@627°F			@ 632°F	
Hydrocarbons	174,648	23,990	32,760	357,138	181	1,976.9
Steam	–	–	–	28,980	18	1,610.0
Total	174,648	23,990	32,760	386,118	107.6	3,586.9
Tray 22 *		@ 498°F		@500°F		
Hydrocarbons	122,412	17,487	22,809	278,148	154	1,802.9
Steam	–			30,818	18	1,027.3
Total	122,412	17,487	22,809	308,966	109	2,830.2
Tray 34		@367°F			@370°F	
Hydrocarbons	21,540	3,254	3,925	133,050	115	1,157.8
Steam				33,968	18	1,887.1
Total	21,540	3,254	3,925	167,018	55	3,044.9
Tray 45		@ 255°F			@ 262°F	
Hydrocarbons	106,438	16,735	19,029	183,089	112.6	1,626.1
Steam				37,381	18	2,076.7
Total	106,438	16,735	19,029	220,469	59.5	3,702.8

*Does not include pumparound liquid stream.

Tower diameter calculations

The calculation for tower vapor flood capacity is given by the expression

$G_f = K_r\sqrt{\rho_v \times (\rho_l - \rho_v)}$
G_f = Mass velocity of vapor at flood in lbs/hr · sqft of bubble area.
K_r = 1,100 for a 24″ tray spacing

The largest diameter will always be below the bottom pumparound in terms of the vapor load. For a quick estimate let the bubble area be 80% of the total tower diameter.

Below bottom pumparound loading

Assume bottom pumparound returns to tower at 300°F.

Total PA duty = 17,740,000 Btu/hr
Enthalpy at 627°F = 347 Btu/lb
Enthalpy at 300°F = 140 Btu/lb

The lbs/hr of the PA stream is $\frac{17{,}740{,}000}{207}$ = 85,715lbs/hr or 11,774 GPH
GPH of unstripped HGO product = 3,668
GPH of overflow = 23,990
Total liquid flow on the tray = 39,632 GPH @ 60°F

The loading data on this tray is as follows:

	Vapor	Liquid
Temperature °F	632	627
Pressure psig	23.8	–
Moles /hr	3,586.9	GPH hot 54,120
lbs/hr	386,118	lbs/hr 288,522
ACFS	302	2.0
ρ_v lbs/cuft	0.355	lbs/cuft 39.9

$$\begin{aligned} \text{Gf} &= 1{,}100\sqrt{0.355 \times (39.9 - 0.355)} \\ &= 4{,}121 \text{ lbs/hr} \cdot \text{sqft.} \\ \text{Ga} &= 4{,}121 \times 0.8 = 3{,}297 \text{ lbs/hr sqft} \end{aligned}$$

Tray area required to handle vapor = $\frac{386{,}118}{3{,}297}$ = 117 sqft
this may be considered as the bubble area.

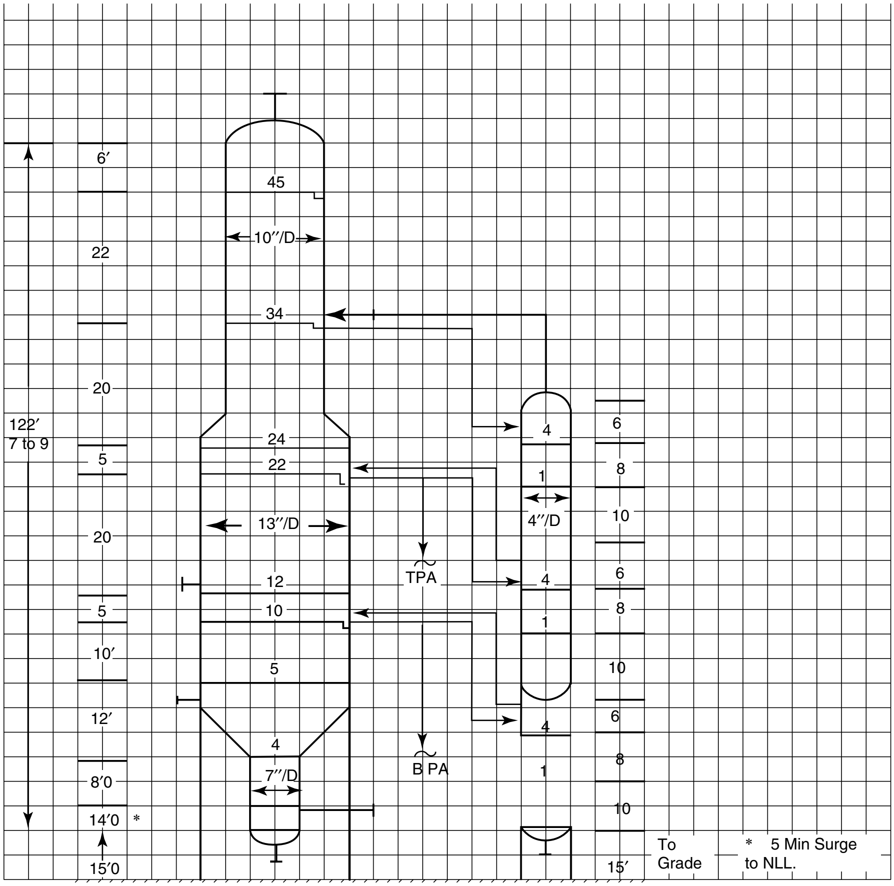


Figure 3.24. Main tower and side stream tower diagram.

Let d'comer area be such that liquid flow is 0.6 ft/sec. Area required is

$$\frac{2.0\text{ACFS}}{0.6} = 3 \text{ sqft}$$

For two (inlet + outlet) $A_{\text{dc}} = 6$ sqft.

Waste area is taken as 20% of $A_{\text{b}} = 117 \times .2 = 23.4$ sqft.

Then $A_{\text{s}} = 117 + 23.4 + 6 = 146$ sqft.

Tower diameter at this location is 13.6 ft.i/d.

Other sections of the tower

The other sections of the tower where there may be changes in diameter are below the top pumparound draw off, and of course the bottoms stripper top tray. The tower top tray should also be checked for loading.

The same calculation is followed for these other sections but it is not proposed to show them here. The results of the calculations though gave the following:

Top section of the main tower above tray 24 diameter is 10 ft i/d
Bottoms stripper section below tray 4 diameter will be 7 ft i/d.

A diagram of the main tower and the associated stripper tower is shown in Figure 3.1.24.

3.2 The vacuum crude distillation unit

As an introduction to this part of the chapter, it will be of interest to outline briefly an important development that occurred in this process during the early 1960s. Originally vacuum units followed closely on design to the atmospheric unit except of course it operated under a vacuum condition. The vacuum was obtained by a two or three stage steam ejectors and the internals of the tower were traditional trays, mostly bubble cap type. Under these conditions the vacuum obtained in the flash zone required the injection of steam to provide the required hydrocarbon partial pressure for adequate vaporization of the fuel oil feed. With the molecular weight of steam low at 18 the tower vapor traffic was extremely high in velocity requiring a large tower diameter to accommodate it.

The break-through to provide vacuum towers of much lower diameters came in the '60s with the use of high capacity steam injectors producing very low vacuum condition in the tower overhead. This coupled with the development of highly efficient expanded grid internals with very low pressure drop allowed the desired flash zone conditions to be met with out the injection of steam. This process became known as the 'Dry Vac' process and is the accepted process now for vacuum crude distillation. Such a process is describes below.

Process description

This process is often integrated with the Atmospheric Crude Distillation unit as far as heat transfer is concerned. Generally the atmospheric residue from the CDU is routed hot to the fired heater of the vacuum unit (Figure 3.25).

The atmospheric residue is further distilled to provide the heavy distillate streams used for producing lube oil or as feed to conversion units. This distillation however has to

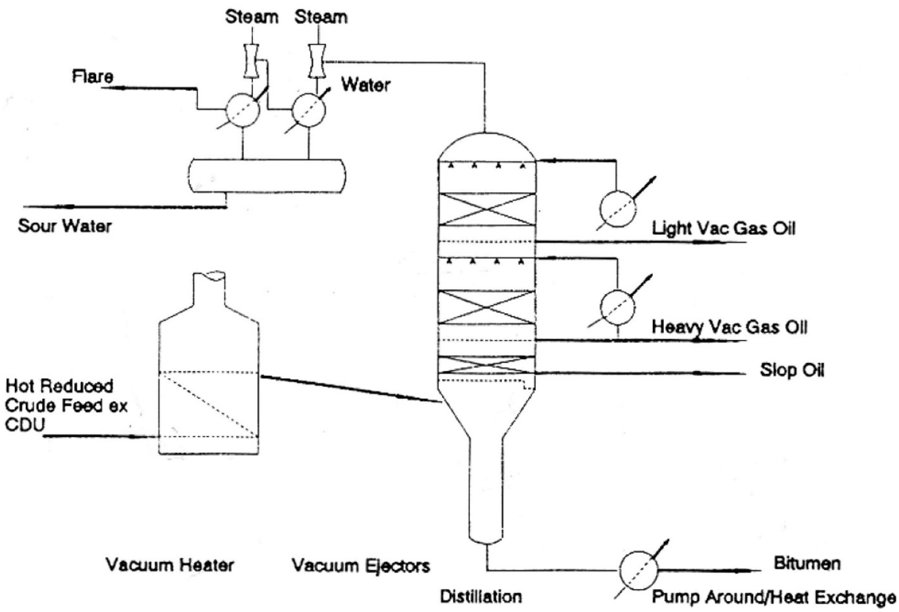


Figure 3.25. Vacuum distillation unit process schematic.

be conducted under sub atmospheric pressure conditions. The temperature required for vaporising the residue at atmospheric pressure would be too high and the crude would crack.

The process follows very much the same pattern as the atmospheric distillation. Should the cold feed be pumped from storage, it is heat exchanged against hot product and pumparound streams before being vaporised in the distillation unit heater. Normally though the feed is pumped hot directly from the CDU's residue stripper to the vacuum unit's heater. Thereafter the distillate vapours are condensed in the tower by heat and mass transfer with the cold reflux streams moving down the tower in the same way as the side streams in the Atmospheric unit. The products are taken off at the appropriate sections are cooled either by heat exchange with colder streams in the atmospheric unit, by air coolers or, in some cases as heating mediums to light end reboilers. They are then pumped to storage.

Neither the vacuum residue that leaves the bottom of the tower in this process nor the side-streams are steam stripped. The vacuum condition is produced by steam ejectors taking suction from the top of the tower. These ejectors remove inert and other vapour that may exist and pull a vacuum of about 5 mmHG absolute. The tower internals are usually expanded grid type which offer low pressure drop such that the flash zone pressure is about 25–30 mmHg absolute.

The vacuum crude distillation unit's flash zone

At atmospheric condition of pressure the flash temperature of normal atmospheric residue feed to achieve any meaningful degree of vaporisation would be extremely high (say in excess of 900°F). At these kind of temperatures the heavy residue will begin to break up or crack. This forms coke in the extreme and olefinic products which may not be desirable to the refiner. Effective vaporisation and fractionation can be achieved however at reduced pressures. Under this condition a reasonable flash temperature (say 650–750°F) can be easily obtained.

As described earlier vacuum distillation units handling reduced crude operate at 3–5 mmHg at the top of the tower and about 25–30 mmHg in the flash zone. No steam is used for stripping. The oil can still crack of course if the cut point desired is so high that excessively high flash temperature is required to meet it even at reduced pressures. The following graph is a guide to the critical cracking temperatures (Figure 3.26).

This graph shows a plot of a range of temperatures within which the oil will begin to crack. This is correlated to the Watson characteristic factor 'K'. Most residuum with a 700°F cut point for Middle East crudes have a 'K' factor of about 11.5. From the

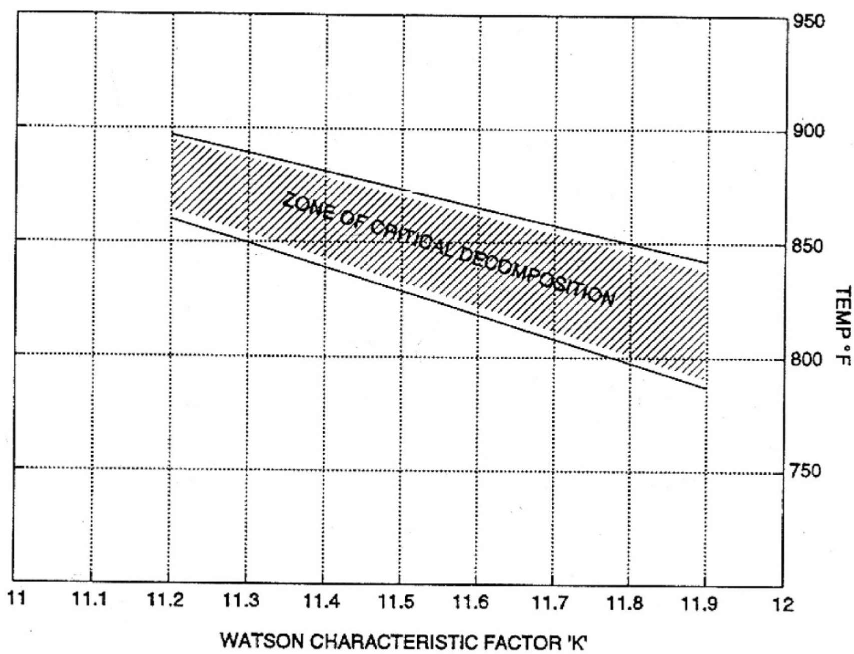


Figure 3.26. Critical cracking zone.

curve therefore it can be seen that these residuum would begin to crack at temperature between 830°F and 855°F. The degree of cracking at or above the 'Zone of critical decomposition' will be a function of temperature and the residence time of the oil at that temperature.

Significant cracking of the oil in a vacuum tower causes:

- High load to the ejectors (due to the formation of light ends)
- In lube oil production, de-colorising of the distillate streams
- In producing feed to hydro-treaters or hydro-crackers—high hydrogen consumption in these units due to the presence of unsaturates as the product of cracking

It is therefore very desirable to avoid these critical temperature in a vacuum unit.

The tower overhead ejector system

Most 'Dry Vac' Towers have a similar overhead ejector system, and as the design of this is critical to the units operation a calculation method to determine this design is described below as calculation steps:

The calculation procedure that is described here relates to the 'Dry' vacuum unit where no steam is used in the distillation process itself. This method can be used to determine the efficiency of the installed ejector set under test run conditions or indeed to specify the equipment to be purchased. The efficiency of the ejectors may be determined by the actual quantities of steam used to that calculated by this method.

The following data needs to be obtained to use this procedure:

- Quantity of inerts—either measured at the exhaust side of the last stage or established as a design criteria
- Tower top temperature and pressure required or observed
- Intermediate stage outlet temperatures and pressures of the process streams
- Total steam flow or steam flow to each stage ejector

A diagram of a typical ejector set is as shown in Fig. 3.27.

The calculation proceeds with the following:

Step 1. Determine the quantity of inerts entering the system from the tower. If this cannot be measured a rule of thumb is that total inerts is 0.5% to 1.0% by weight on feed. This is made up of air leaking into the system and some light ends. Again by rule of thumb light ends will be about 25% of total inerts.

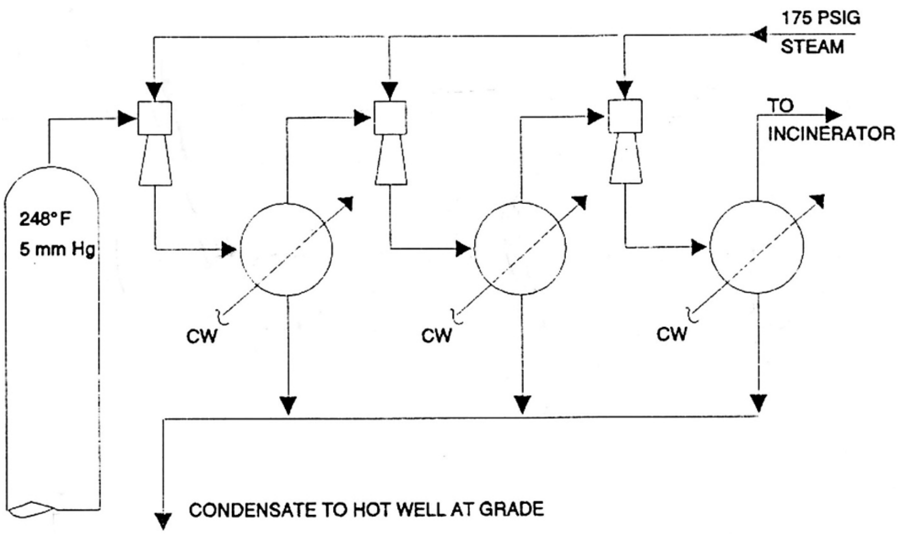


Figure 3.27. Typical ejector set.

Step 2. Calculate the 'Equivalent Dry Air Load' to the first ejector stage. Using the equation:

$$W_a = \frac{W_i}{Rm_i \times Rt_i} + \frac{W_s}{Rm_s \times Rt_s}$$

where

W_a = Equivalent air flow in lbs/hr.
Wi = Actual lbs/hr of o/head (includes air, and light ends).
Rm$_i$ and Rt$_i$ = Ratio factors for component i from ejector Figures 3.28 and 3.29
W_s = Weight flow of steam in lbs/hr.
Rm_s and Rt_s = Ratio factors for steam from Figures 3.30 and 3.31.

In this case W_s will be zero, as no steam will be used in the distillation.

Step 3. Calculate the steam consumption to the first ejector using tower top pressure as the suction pressure. The consumption is calculated by

$$W_{ms} = R_a M_p W_a$$

where

W_{ms} = Weight flow of motive steam in lbs/hr.
R_a = Ratio of lb motive steam/lb air equivalent using Figure 3.30
M_r = Steam usage multiplier from Figure 3.31
W_a = Air equivalent flow in lbs/hr.

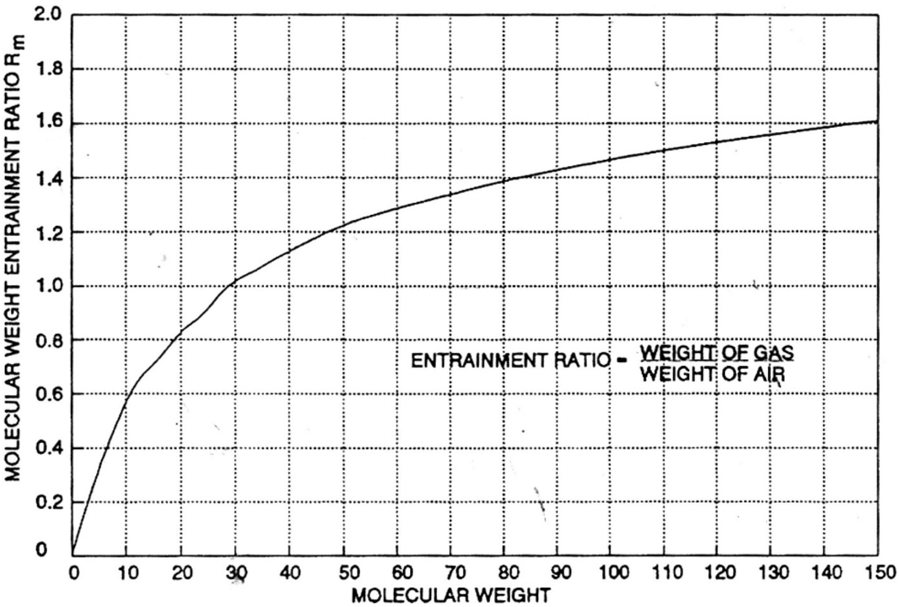


Figure 3.28. Ratio factors for steam.

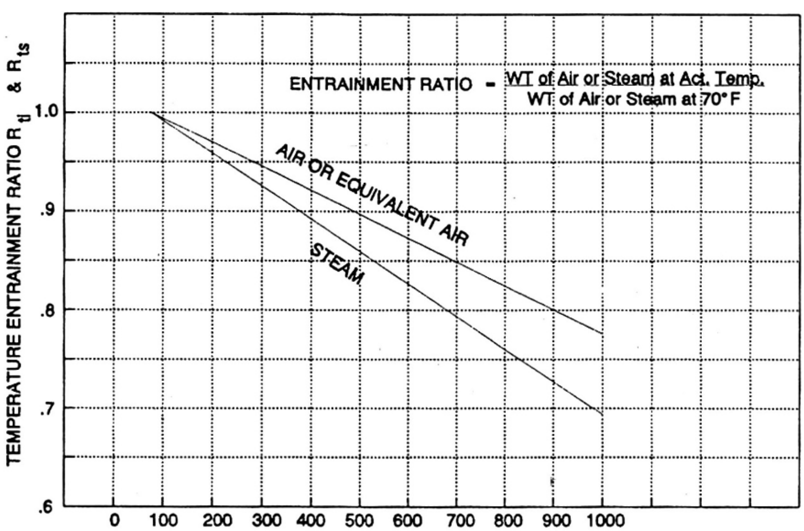


Figure 3.29. Temperature entrainment ratios Rt_i, Rt_s.

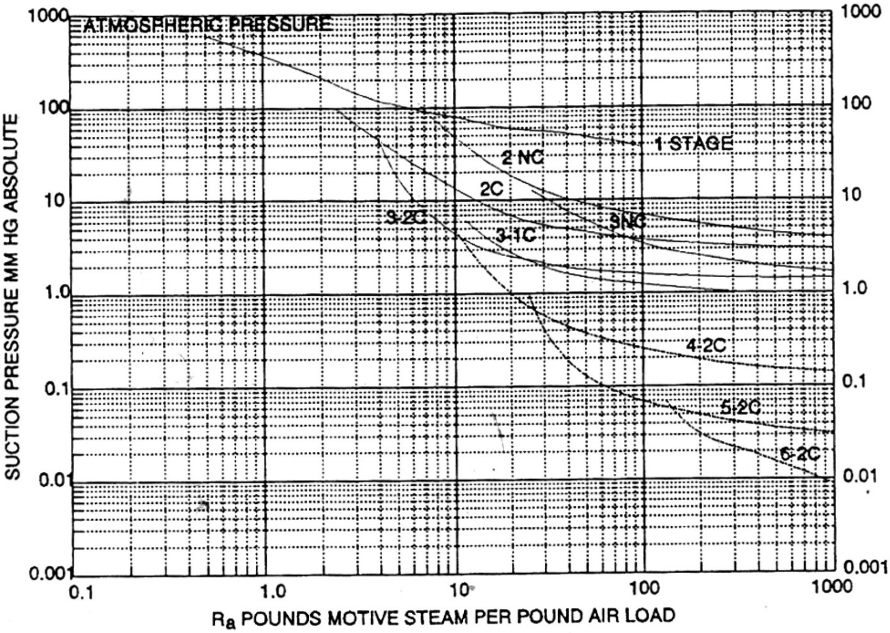


Figure 3.30. Suction pressure V's weight of motive steam.

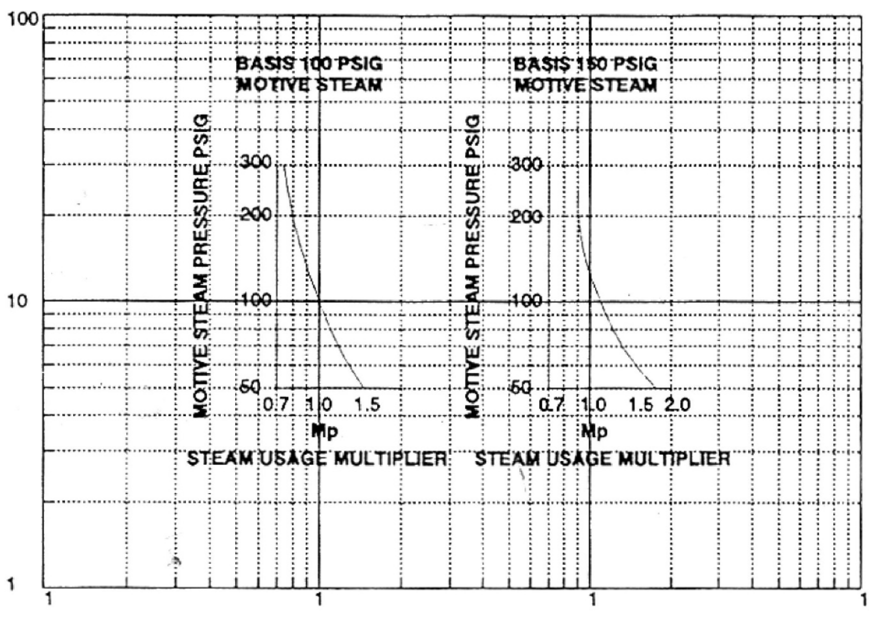


Figure 3.31. Motive steam usage correction factor.

Step 4. Calculate the partial pressure of steam at the condenser pressure. This pressure is the 1st/2nd stage intermediate pressure and is the suction pressure to the 2nd ejector. It is read on the plant or it may be assumed. In the calculation its assumed to be 50 mmHg.

Step 5. From the partial pressure of steam the condensing temperature in the condenser is read from steam tables. Assume 90% steam is condensed in this first condenser.

Step 6. (Optional) Calculate heat balance across the condenser and arrive at the condenser duty.

Step 7. Repeat step 2 for 'equivalent dry air load' to stage 2. Note there will now be steam present in this calculation.

Step 8. Repeat steps 3–5 for the second stage, making an assumption for the condenser pressure or reading off the actual pressure if the unit is an installed one. Assume also the amount of steam condensed in the second condenser—this will be high around 98%.

Step 9. Repeat steps 6 and 7 for the third stage condenser and ejector. This condenser will be about 1 or 2 psi above atmospheric pressure.

Step 10. Summarize the above results into a process specification for a required ejector set or if already installed compare the performance (e.g. calculated steam consumed versus actual).

Calculating flash zone conditions in a vacuum unit

Flash zone conditions are easier to calculate for a 'Dry' vacuum unit than for the atmospheric crude unit. Indeed the flash zone conditions can be measured in vacuum units with a greater degree of accuracy than in the case of the atmospheric column.

The procedure for predicting flash zone conditions in this case follows a similar route to that in the atmospheric unit case. The following steps describe this procedure.

Step 1. Develop the EFV from the TBP curve of the reduced crude. The same method that was used in item discussing the flash zone of the Atmospheric Crude Unit in this chapter will apply in this case also. Remember the EFV calculated is for atmospheric pressure.

Step 2. Develop the material balance for the vacuum unit. As the flash zone conditions are to be used in most calculations involving the vacuum tower it is best to develop the materials balance at this stage. To do this determine the distillate cuts required and by using the respective crude assay determine the specific gravity for each cut using mid boiling points. The mid boiling point for each distillate product only should be determined in this case. Use the method described in Chapter 1 of this Handbook to determine mol weight. Some of these will need to be extrapolated.

Step 3. Set the overflash. Now in this area of the TBP and EFV curves the slope of the curve is quite steep. That is there is a high temperature difference for each percent of volume increase. In vacuum units therefore a 1–2% over flash would be realistic to meet a realistic coil outlet temperature.

Step 4. Determine the new flash temperature to include the over flash from the EFV curve. This is the temperature at atmospheric pressure.

Step 5. Calculate the total pressure of the flash zone. Vacuum towers operate at or below 5 mmHg at the top. Pressure drop through the tower should not be more than 25 mmHg. A well designed off take trays and packing should be as follows:

Grid packing 6–7 mmHg per 10 foot of packed height.
Draw off (chimney trays) 2–3 mmHg/tray.

Step 6. There will be no partial pressure calculation of hydrocarbon vapor (as in the case of the atmospheric unit) as there is no steam in the flash zone of a 'Dry' vacuum tower. The total pressure calculated in Step 5 is the actual hydrocarbon flash pressure. Using the vapor pressure curves determine the flash zone temperature at the total flash zone pressure. This is the flash zone temperature that will now be used for all heat balances, etc.

Draw-off temperatures

Unlike the atmospheric crude distillation unit the temperature of the vacuum tower bottom (Bitumen) will be essentially the flash zone temperature. There will be a small difference, say 2–3°F, below actual flash zone temperature due to over flash returning from the wash trays. Very often the overflash amount is drawn off from below the wash section and either sent to fuel or blended into the bitumen stream external to the tower. In this case the unquenched bitumen leaving the tower will be at flash zone temperature. Again there is no steam present to influence this temperature.

Side stream draw-off temperatures are easier to calculate for a vacuum tower than was the case for the atmospheric tower. This is so because in a 'dry vacuum' column there is no steam to influence partial pressures and of course there is no side stream stripping.

A method similar to the 'Packie' method used for the atmospheric column is used for the vacuum column draw off. In this case however it is only necessary to determine the initial boiling point of the side stream EFV curve at the tower condition to arrive at the draw-off temperature.

Note: It is the IBP of the actual EFV curve in this case NOT the IBP of the flash reference line as in the case of the atmospheric unit. Also there will be no 'Packie' correction factors required in this case.

The calculation steps for this procedure are as follows:

Step 1. Draw the EFV curve from the sides tream TBP curve using the method described earlier in this chapter. Only the 0, 10, 30, and 50% vol section of the curve need to be developed.

Step 2. Set the total pressure at the draw-off tray. If this is not available as plant data then use the criteria for pressure drop given in the item dealing with the flash zone.

Step 3. Calculate the partial pressure of the side stream product at the draw-off tray. To do this consider all material lighter than the draw off side stream to be inert. Include in the inert estimates of air leakage and cracked hydrocarbon vapors as described earlier in this chapter in the item dealing with *Tower Overhead Ejector System*. The total hydrocarbon vapor will include the overflow from the draw-off tray. As a rule of thumb estimate overflow as:

Top side stream	0.8 times product
Mid side stream	1.0 times product
Bottom side stream	1.5 to 2.0 times product

Step 4. Using the vapor pressure curves relate the IBP temperature of the EFV to the partial pressure determined in Step 3. This is the draw-off temperature, and this will be the temperature for the respective side streams and pumparound draw off that will be used in the tower heat balances.

Determine pumparound and internal flows for vacuum towers

Now that the cut points and tower conditions of temperature and pressure are established the internal flow and pumparound duties can be calculated, although generally speaking fractionation requirements are not as strict in a vacuum crude unit as in the case for the atmospheric unit. Nevertheless proper wash streams are required in vacuum towers to protect distillates which nay become feed to cracking units, from entrained undesirable components such as metals. Test runs on vacuum units therefore should include the determination of reflux streams and, in turn, tower loading.

The following steps outline a calculation procedure to determine pumparound requirements and overflow (reflux) in the wash section of the tower.

Step 1. Set the overflow requirement for the LVGO draw-off tray using the rule of thumb given in the previous item. Alternatively if this can be measured on the plant use that data.

Step 2. From plant data or such data as can be developed from items on ejector system and draw-off temperatures, calculate the heat balance below the LVGO draw-off tray.

Step 3. In this heat balance the bottom pumparound duty will be the unknown. Equate heat in equals heat out to determine the duty of the pumparound required to produce the set overflow.

Step 4. This pumparound duty can be checked on the plant by multiplying the flow in the pumparound by the enthalpy difference over the exchangers.

Step 5. Carry out the overall heat balance over the tower. That is, calculate the difference between the total heat in with the feed and the total out with all the products. This difference gives the total heat to be removed by both pumparounds. Assuming there are two pumparounds (top and bottom). The duty of the bottom pumparound has already been calculated. Then the top pumparound duty will be the total heat to be removed minus the duty of the bottom pumparound.

Step 6. Usually the most critical flow in a vacuum unit is the wash oil flowing over the bottom wash trays or packing. This is the area where most undesirable entrainment can occur and this is the most vulnerable area for coking. Lack of wash oil enhances contamination of the bottom product and promotes coking in this area.

Step 7. Carry out a heat balance over the bottom wash section of the tower. The unknown in this case is the overflow liquid from the heavy vacuum gas oil. Equate the heat in with feed and overflow with the heat out with total product vapors, overflow vapor, and bitumen to solve for the unknown.

Note: The quantity of overflow in this case is independent of pumparound duties above it. It is dependant only on the amount of over flash.

Calculate tower loading in the packed section of vacuum towers

As discussed earlier, most modern 'dry' vacuum towers use low pressure drop grid or stacked packing. This packing enhances heat exchange in the tower and of course permits the tower to operate at very low pressures. Nevertheless this packing can become overloaded causing high pressure drop in the tower and poor all-around performance.

This item describes a general method of evaluating the grid performance in terms of its pressure drop. Please note this is a quick general method of estimating tower packing design or performance. Proprietary grid and packing manufacturers have their own correlations which they use in their design work. A more detailed examination of packed tower loading is presented in the author's published work titled '*Elements of Chemical Process Engineering'**. The following are the calculation steps used for this quick general packed section evaluation.

Step 1. Determine the liquid and vapor flows across the section to be evaluated. As calculated in the previous section "determine pumparound and internal flows for vacuum towers."

Step 2. If the unit is existing and this calculation is to determine tower performance then use manufacturers drawings for tower details such as dimensions of the packed section.

Step 3. Calculate the liquid and vapor loads in terms of actual cubic feet per second for vapor and cubic feet per hour per square foot of tower for the liquid. All these will be at tray conditions of temperature and pressure.

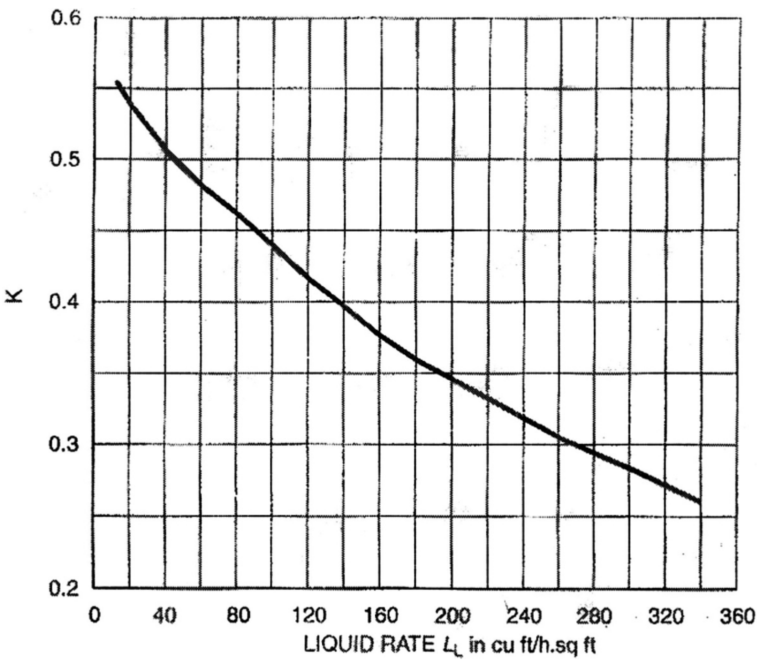


Figure 3.32. Capacity factor 'K' V's liquid rate.

Step 4. Using the liquid load as calculated in step 3, read off the value for 'K' from Figure 3.32.

Calculate the linear velocity of the vapor at flood from the expression

$$K = V_f \sqrt{\rho_v / (\rho_l - \rho_v)}$$

where:

V_f = Vapor velocity at flood in ft/sec.
ρ_v = Density of vapor in lbs/cuft at section conditions of temperature and pressure
ρ_l = Density of liquid in lbs/cuft at section conditions.

Step 5. Calculate the actual vapor velocity required by multiplying the calculated velocity at flood by percent of flood permissible if this is to be a new design. The design cross sectional area of the tower is then the calculated vapor load in cuft/sec divided by the actual vapor velocity. If the unit is existing divide the vapor loading in cuft/sec by the cross sectional area of the tower to arrive at the actual vapor velocity in ft/sec. The existing unit operation as a percent of flood will be the actual velocity divided by the calculated flood velocity times 100.

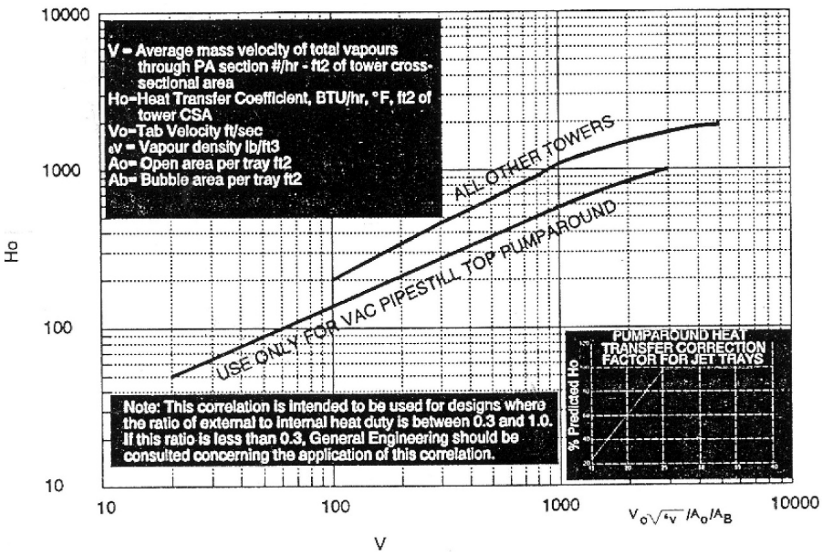


Figure 3.33. Transfer coefficient H_o V's mass velocity for pumparound zones.

Step 6. Estimate the HETP (height equivalent to a theoretical tray) as being between 1.5 and 2.0 ft. Use the higher figure for vapor percent of flood 50–95% or higher and the lower value for vapor flood below 50%.

Step 7. Calculate the height of the packed section required for heat transfer using the quantity of heat to be transferred as being the pumparound duty in Btu/hr. Then treat the section above the pumparound draw off as a simple heat exchanger. Use the liquid flow to this section and the pumparound flow as calculated in the previous section (as part of the tower liquid load). The inlet temperature to the section can be taken as that for the pumparound liquid flow into the tower and the temperature out of the section as the pumparound draw-off temperature.

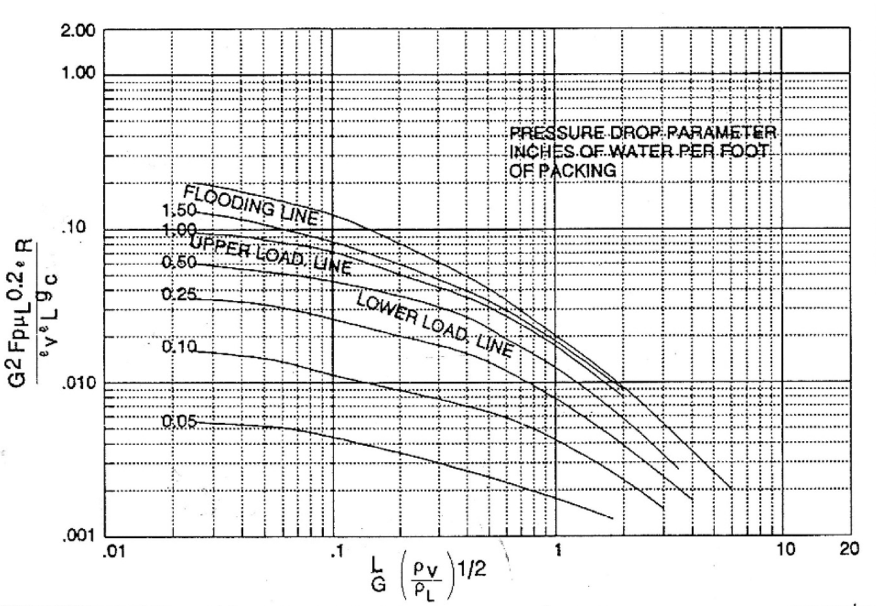
Step 8. Read off an overall heat transfer coefficient Ho from Figure 3.33.

Calculate the LMTD over the section from the temperatures used in the previous section on pumparound etc. Then calculate the total area of tower required for the heat transfer from the expression:

$$Q = AH_o\Delta t_m$$

where

Q = Heat duty in Btu/hr
A = Heat transfer area in sqft
H_o = Overall heat transfer coefficient in Btu/sqft·hr·°F
Δt_m = Log mean temperature difference in °F.



L	= Liquid Rate, Lbs/sec., Sq.ft	G	= Gas Rate, Lbs/sec., sq.ft
e_L	= Liquid Density, Lbs/cu. ft	e_V	= Gas Density, Lbs/cu. ft
F_p	= Packing Factor $\approx a/\in^3$	μ_L	= Viscosity of Liquid, Centipoise
e_R	= Ratio, $\frac{\text{Density of Water}}{\text{Density of Liquid}}$	g_c	= Gravitational Conversion Factor = 32.2

Figure 3.34. Pressure drop through grid in inches of hot liquid per foot height.

Step 9. Calculate the theoretical number of trays required by dividing the total area calculated in step 8 by the design cross sectional area (step 5). Multiply these number of trays by the selected HETP to give the height of packing.

Step 10. Calculate the pressure drop through the grid using the actual vapor velocity in cuft/sec in the equation $K = V_a\sqrt{\rho_v(\rho_l - \rho_v)}$ to determine the constant K the read off the pressure drop in inches of hot liquid per foot height from Figure 3.34.

To express this pressure drop in mmHg multiply by the SG of the hot liquid and 1.865.

Appendix

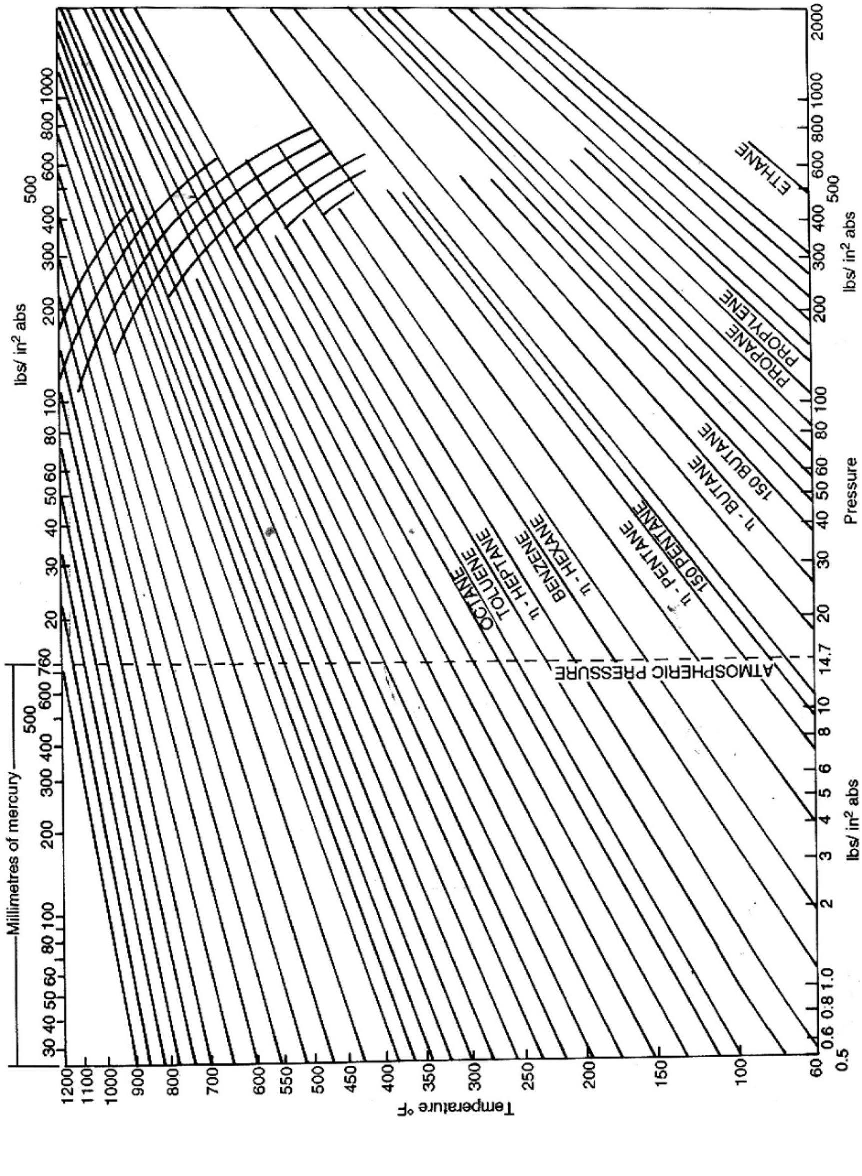


Figure 3.A.1. Pressure temperature curves (2 Pages).

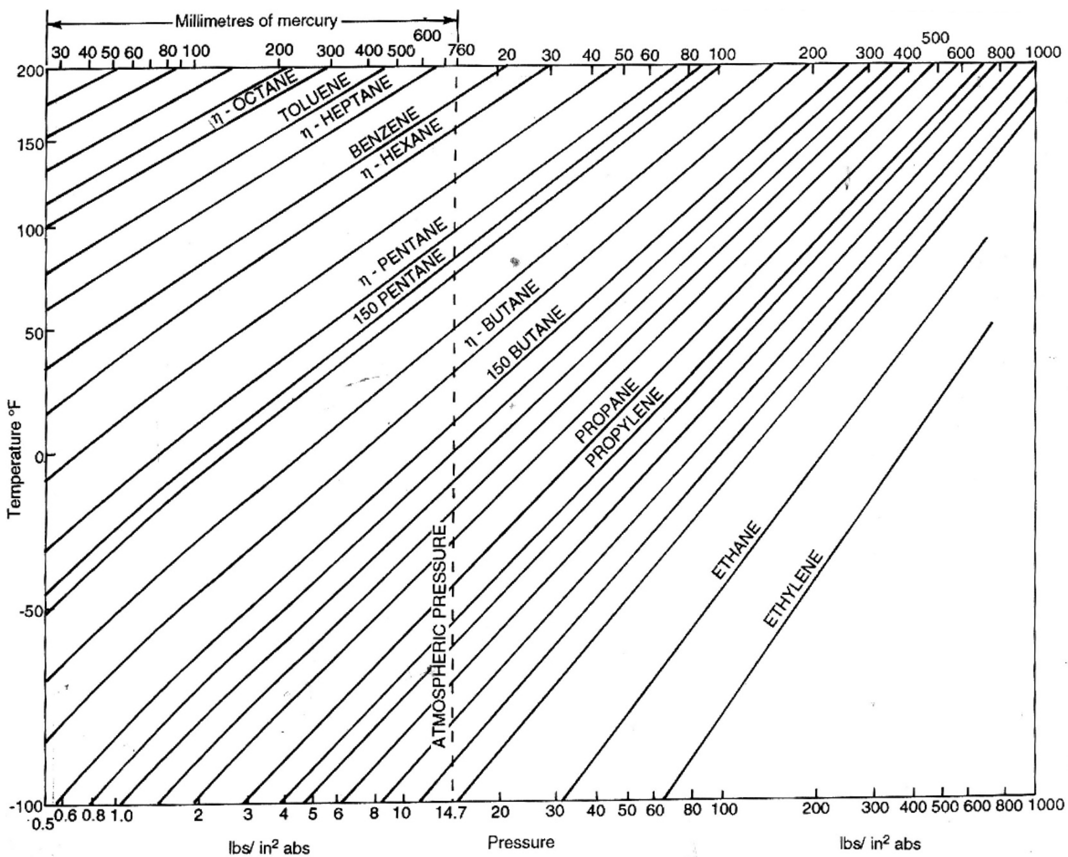
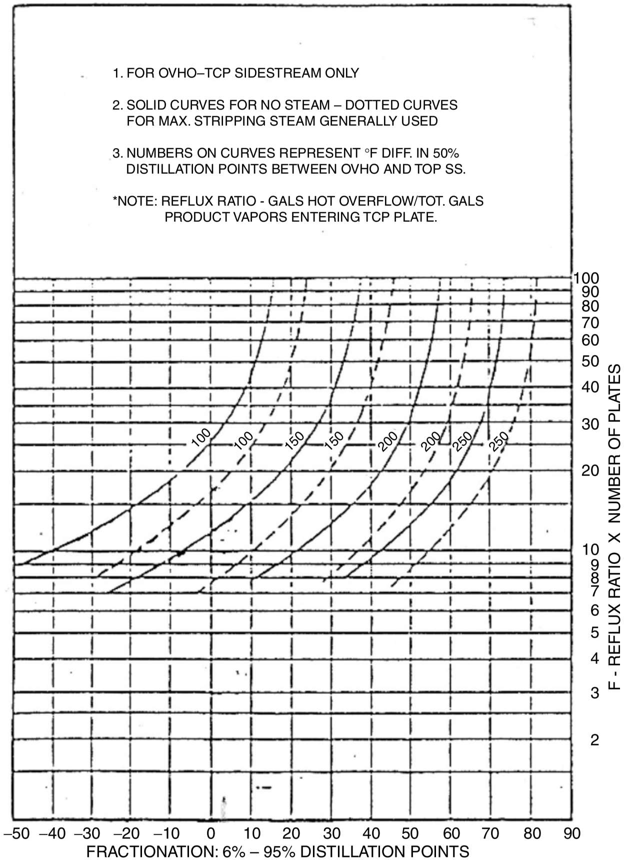
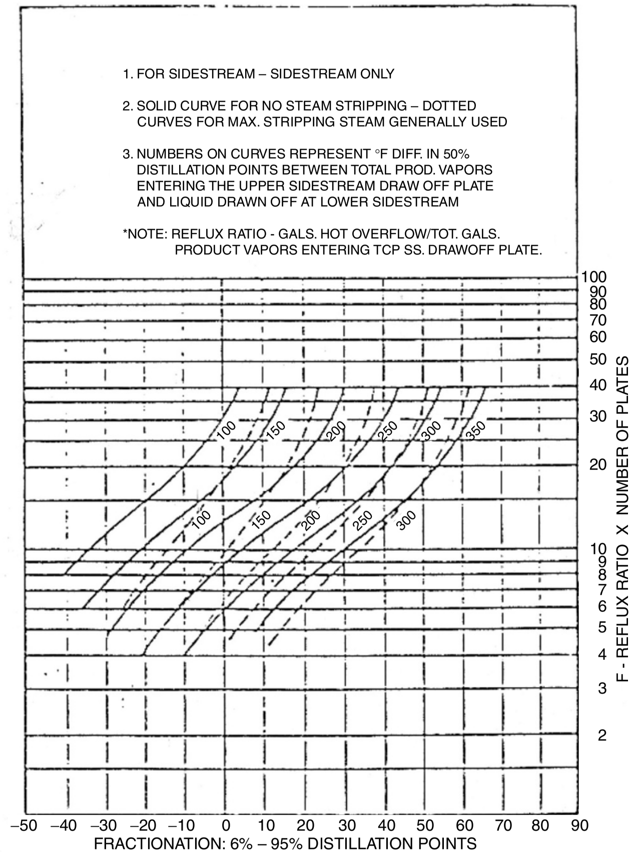


Figure 3.A.1. (Cont.)



Fractionation Curves for Overhead to Top Sidestream Products.



Fractionation Curves for Sidestream to Sidestream Products.

Figure 3.A.2. ASTM gaps and overlaps.

FIGURE A 3.0 VALVE TRAY DESIGN PRINCIPLES			
Design Feature	**Suggested Value**	**Alternate Values**	**Comment**
1. **Valve Size and Layout**			
a. Valve diameter	–		Valve diameter is fixed by the vendor
b. Percent Hole Area, A_O/A_B	12	8 to 15	Open area should be set by the designer. In general, the lower and open area, the higher the efficiency and flexibility, and the lower the capacity (due to increased pressure drop). At values of open area toward the upper end of the range (say 15%), the flexibility and efficiency are approaching sieve tray values. At the lower end of the range, capacity and downcomer filling becomes limited
c. Valve Pitch/diam. ratio	–		Valve pitch is normally triangular. However, this variable is usually fixed by the vendor
d. Valve distribution-		-	On trays with flow path length ≥ 5', and for liquid rates > 5000 GPH/ft. (diameter) on trays with flow path length < 5', provide 10% more valves on the inlet half of the tray than on the outlet half
e. Bubble Area, A_B	–		Bubble area should be maximised
f. Plate efficiency	–		Valve tray efficiency will be about equal to sieve tray efficiency provided there is not a blowing or flooding limitation
g. Valve blanking	–		This should not generally be necessary unless tower is being sized for future service at much higher rates. Blanking strips can then be used. Blank within bubble area, not around periphery to maintain best efficiency
2. **Tray Spacing, Inches**	-	12 to 36	Generally economic to use min. values given on p. III-E-2 which are set by maintenance requirements. Other considerations are downcomer filling and flexibility. Use of variable spacings to accommodate loading changes from section to section should be considered.
3. **Number of Liquid Passes**	11	to 2M	ultipassing improves liquid handling capacity at the expense of vapour capacity for a given diameter column and tray spacing. Cost is apparently no greater - at least, for tower diameters < 8 ft.
4. **Downcomers and Weirs**			
a. Allowable Downcomer inlet velocity, ft/sec of clear liq.		0.3 to 0.4	Lower value recommended for absorbers or other systems of known high frothiness
b. Type downcomer	Chord	Chord, Arc	Min. chord length should be 65% of tray diameter for good liquid distribution. Sloped downcomers can be used for high liquid rates - with maximum outlet velocity = 0.6 ft/sec. Arc downcomers may be used alternatively to give more bubble area (and higher capacity) but are somewhat more expensive. Min. width should be 6 in. for latter
c. Inboard Downcomer Width (Inlet and Outlet)		Min. 8 inches	Use of a 14-16" "jump baffle" suspended lengthwise in the centre of the inboard downcomer and extending the length of the downcomer is suggested to prevent possible bridging over by froth entering the downcomer from opposite sides. Elevation of base of jump baffle should be level with outlet weirs. Internal accessway must be provided to allow passage from one side to another during inspection
d. Outlet Weir Height	2"	1" to 4"	Weir height can be varied with liquid rate to give a total liquid head on the tray (h_c) in the range of 2.5" - 4" whenever possible. Lower values suggested for vacuum towers, higher ones for long residence time applications
e. Clearance under downcomer, in.	1.5"	1" min	Set clearance to give head loss of approximately 1 inch. Higher values can be used if necessary to assure sealing of downcomer
f. Downcomer Seal (Inlet or outlet weir height minus downcomer clearance)	Use outlet weir to give min. _" seal in plate liquid	Inlet weir or recessed inlet box	In most cases plate liquid level can be made high enough to seal the downcomer through use of outlet weir only. Inlet weirs add to downcomer build up; in some cases they may be desirable for 2-pass trays to ensure equal liquid distribution. Recessed inlets are more expensive but may be necessary in cases where an operating seal would require an excessively high outlet weir
g. Downcomer filling, % of tray spacing		40 to 50	Use the lower value for high pressure towers, absorbers, vacuum towers, known foaming systems, and also for tray spacings of 18" or lower

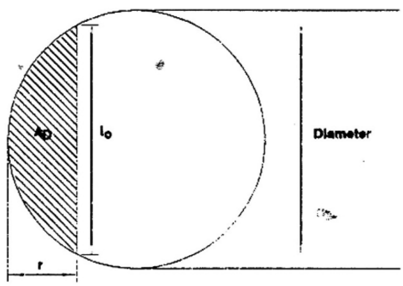
Figure 3.A.3. Valve tray design details.

WEIR LENGTH AND DOWNCOMER AREA

R*	L*	A*	R*	L*	A*	R*	L*	A*
.070	.511	.0308	.120	.650	.0680	.170	.751	.113
1	.514	.0315	1	.652	.0688	1	.753	.114
2	.517	.0321	2	.654	.0697	2	.755	.115
3	.521	.0328	3	.657	.0705	3	.756	.116
4	.524	.0335	4	.659	.0714	4	.758	.117
5	.527	.0342	5	.661	.0722	5	.760	.117
6	.530	.0348	6	.663	.0731	6	.762	.118
7	.533	.0355	7	.665	.0739	7	.763	.119
8	.536	.0362	8	.668	.0748	8	.765	.120
9	.539	.0368	9	.670	.0756	9	.766	.121
.080	.542	.0375	.130	.672	.0765	.180	.768	.122
1	.545	.0382	1	.674	.0774	1	.770	.123
2	.548	.0389	2	.677	.0782	2	.772	.124
3	.552	.0396	3	.679	.0791	3	.773	.125
4	.555	.0403	4	.682	.0799	4	.775	.126
5	.558	.0410	5	.684	.0808	5	.777	.127
6	.561	.0418	6	.686	.0817	6	.778	.128
7	.564	.0425	7	.688	.0825	7	.780	.129
8	.567	.0432	8	.691	.0834	8	.781	.130
9	.570	.0439	9	.693	.0842	9	.783	.131
.090	.573	.0446	.140	.695	.0851	.190	.784	.132
1	.576	.0454	1	.697	.0860	1	.786	.133
2	.578	.0461	2	.699	.0869	2	.787	.134
3	.581	.0469	3	.700	.0878	3	.789	.135
4	.583	.0476	4	.702	.0887	4	.790	.136
5	.586	.0484	5	.704	.0896	5	.792	.137
6	.589	.0491	6	.706	.0905	6	.794	.138
7	.592	.0499	7	.708	.0914	7	.795	.139
8	.594	.0506	8	.710	.0923	8	.797	.140
9	.597	.0514	9	.712	.0932	9	.798	.141
.100	.600	.0521	.150	.714	.0941	.200	.800	.142
1	.603	.0529	1	.716	.0950	1	.802	.143
2	.605	.0537	2	.718	.0959	2	.803	.144
3	.608	.0545	3	.720	.0969	3	.805	.145
4	.610	.0553	4	.722	.0978	4	.806	.146
5	.613	.0561	5	.724	.0987	5	.808	.148
6	.615	.0568	6	.726	.0996	6	.809	.149
7	.618	.0576	7	.728	.1005	7	.810	.150
8	.620	.0584	8	.729	.1015	8	.812	.151
9	.623	.0592	9	.731	.102	9	.813	.152
.110	.625	.0600	.160	.733	.103	.210	.814	.153
1	.628	.0608	1	.735	.104	1	.816	.154
2	.630	.0616	2	.737	.105	2	.817	.155
3	.633	.0624	3	.738	.106	3	.819	.156
4	.635	.0632	4	.740	.107	4	.820	.157
5	.638	.0640	5	.742	.108	5	.822	.158
6	.640	.0648	6	.744	.109	6	.823	.159
7	.643	.0656	7	.746	.110	7	.824	.160
8	.645	.0664	6	.747	.111	8	.826	.161
9	.648	.0672	9	.749	.112	9	.627	.162

R*	L*	A*	R*	L*	A*	R*	L*	A*
.220	.828	.163	.280	.898	.230	.390	.977	.361
1	.829	.164	5	.903	.236	5	.979	.367
2	.831	.165						
3	.832	.166	.290	.908	.241	.400	.980	.374
4	.834	.167	5	.913	.247	5	.982	.380
5	.835	.169	.300	.917	.252	.410	.984	.386
6	.836	.170	5	.921	.258	5	.986	.392
7	.838	.171						
8	.839	.172	.310	.925	.264	.420	.987	.398
9	.841	.173	5	.930	.270	5	.989	.405
.230	.842	.174	.320	.933	.276	.430	.991	.412
1	.843	.175	5	.937	.282	5	.993	.418
2	.845	.176						
3	.846	.177	.330	.941	.288	.440	.994	.424
4	.848	.178	5	.945	.294	5	.995	.430
5	.849	.179	.340	.948	.300	.450	.996	.437
6	.850	.180	5	.951	.306			
7	.851	.181				.460	.997	.450
8	.853	.182	.350	.955	.312			
9	.854	.183	5	.958	.318	.470	.998	.462
.240	.855	.184	.360	.961	.324	.480	.998	.475
5	.860	.190	5	.964	.330			
						.490	.999	.488
.250	.866	.196	.370	.967	.337			
5	.872	.202	5	.969	.343	.500	1.0	0.50
.260	.878	.207	.380	.971	.348			
5	.883	.213	5	.977	.354			
.270	.888	.218						
5	.893	.224						

* THIS TABLE RELATES THE DOWNCOMER AREA, THE WEIR LENGTH, AND THE HEIGHT OF THE CIRCULAR SEGMENT FORMED BY THE WEIR.



$R = \frac{\text{* Downcomer Rise}}{\text{Diameter}} = \frac{r}{\text{Dia.}}$ $\quad L = \frac{\text{* Weir Length}}{\text{Diameter}} = \frac{lo}{\text{Dia.}}$ $\quad A = \frac{\text{* Downcomer Area}}{\text{Tower Area}} = \frac{A_D}{A_S}$

Figure 3.A.4. Chord height, area, and lengths.